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GP-1 **Technology and Business Information for the Global Gas Processing Industry**

Cover Image: An engineer works with virtual reality. Photo courtesy of Siemens.

HP Industry Perspectives

Capital investments/major trends affecting the global HPI

In late January, Tyler Campbell, Associate Editor, *Hydrocarbon Processing*, *Gas Processing & LNG*, and *H2Tech*, presented an update on the global capital construction landscape at the Engineering and Construction Contracting (ECC) Extra Conference in New Orleans, Louisiana (U.S.). His talk focused on capital projects by region, major trends affecting the global refining, petrochemicals and gas processing/LNG industries, and new project announcements, among several other industry trends.

Active projects. Gulf Energy Information's Global Energy Infrastructure database is tracking more than 1,000 active capital projects in the global hydrocarbon processing industry (**FIG. 1**). Approximately 40% of these projects are within the Asia-Pacific region, followed Eastern Europe, Russia and the CIS, the U.S. and the Middle East.

Active capital projects account for more than \$1.8 T in active capital investments. At nearly \$750 B, Asia accounts for approximately 40% of capital investments globally. The region is followed by the U.S. (\$320 B) and the Middle East (\$280 B). These three regions comprise more than 70% of active capital investments.

Major trends. Some of the major trends highlighted during Nichols' presentation included the continued move of the global refining industry toward lower sulfur content in transportation fuels, as well as the shift towards more use of biofeedstocks.

Robust and increasing petrochemicals demand will continue to spur additional capacity builds, especially in developing nations in the Asia-Pacific region.

As countries continue to increase initiatives to curb emissions, natural gas usage and trade (e.g., LNG) will continue to increase significantly. Demand countries (e.g., China, India) are investing heavily in new import/regasification capacity, while supply countries (e.g., Australia, Qatar and the U.S.) will continue to buildout liquefaction/export capacity to satisfy demand. HP

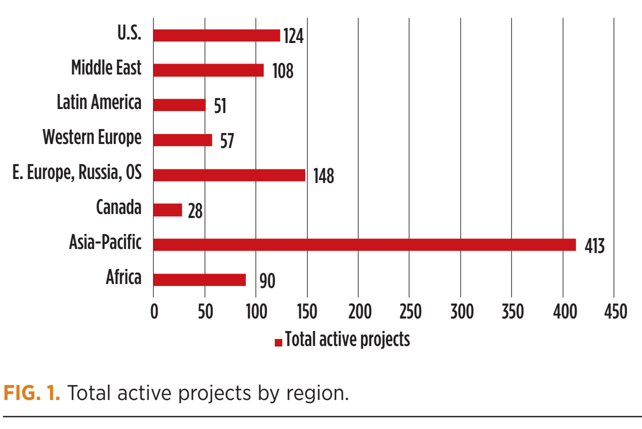

FIG. 1. Total active projects by region.

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Adoption of digital technologies is a mainstay in the HPI

A few years ago, *Hydrocarbon Processing* started to include digital technologies in its monthly article focus. For decades, the publication focused on advancements in plant automation, instrumentation and process controls; however, the industry's digital transformation opened new possibilities and technologies to optimize maintenance, operations, safety, reliability, sustainability and profitability.

Digital transformation within the refining and petrochemical industries has significantly changed not only how plants operate but also how they are design and constructed. The increasing adoption of digital technologies made it imperative to showcase how these advancements are enabling refiners and petrochemical producers to reduce downtime, increase reliability, mitigate emissions and enhance product specifications.

Digital technologies.

This month's issue focuses on the latest digital technologies being adopted, as well as ways to protect systems from cyber attacks. These systems are not intended to replace personnel but to enhance workflows and increase operator efficiencies. The following provides a glimpse of the technologies showcased within this issue.

Molecular modeling. Refiners are facing unprecedented challenges in maintaining their profit margins. Global gasoline demand is forecast to decrease due to the adoption of new fuels and more stringent government regulations are being instituted around the world. A profitable path for refiners is to shift to crude-to-chemicals production. The article "Optimize refining operations using plant digital twin based on molecular modeling" details how molecule-based process simulation can help address the challenges of seeing crude-to-chemicals process benefits, developing refining processes to process alternative feedstocks and more.

Optimizing blending using software. To help find the optimum product mix at minimum costs, refiners typically blend crude oils. To increase margins, refineries look for ways to co-process heavy crude oils with light ones. However, heavier crude oils tend to contain more contaminants that must be processed out or they can damage crucial processing equipment.

The article "Web-based software for predicting crude computability and optimization for increasing heavy oil processing" details a novel method to predict crude oil blending compatibility using prediction model software quickly and effectively.

Robotic process automation. The use of bots, or robotic process automation, is significantly increasing within the global oil and gas supply chain and enabling refined and petrochemical product producers to increase efficiency, productivity, reliability and predictability. The benefits of robotic process automation include automating tedious, time-consuming tasks and error-prone manual processes, freeing personnel to focus on more value-adding work.

Protecting against cyber attacks.

With the increased adoption of digital technologies, the refining and petrochemical industries open themselves to increased cyber attacks.

This issue's Business Trends section details how industry professionals can mitigate downtime and protect their crucial assets to ensure safe and profitable operations.

The refining and petrochemical industries are enhanced with the adoption of new digital technologies, which is why *Hydrocarbon Processing* will continue to expand its coverage to the global processing industries. HP

INSIDE THIS ISSUE

19 Executive Viewpoint. *Hydrocarbon Processing* sat down with Helion Sardinia, Chief Commercial Officer, Lummus Technology, to discuss the state of the hydrocarbon processing industry, key drivers and technologies that are shaping the market, and how the process licensor must evolve for the future.

22 Digital transformation. The adoption of new digital technologies are enabling producers to operate more efficiently, safely and more profitably. This month's Special Focus section details several areas where digital transformation is having significant impacts.

35 Process Optimization. Extractive distillation is one of the most efficient techniques to separate aromatic and non-aromatic hydrocarbons by increasing the relative volatility of the mixture in the presence of a solvent. In this article, Reliance Industries Ltd. details a case study of high pressure drop observed across the extractive distillation column, operational reasons for it and a cost-effective solution to this challenge.

52 History of the HPI. *Hydrocarbon Processing* continues its series on the evolution of the global refining and petrochemicals industries. This installment details the major inventions and technologies developed during the 1930s. This includes the discovery of catalytic cracking, polyethylene, several new chemical products and the use of the first jet-powered aircraft. The section also features early industry pioneers, and a mixture of technical articles, columns and headlines published in the 1930s by the forerunner of *Hydrocarbon Processing, The Refiner and Natural Gasoline Manufacturer.*

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Protecting OT environments against targeted cyber attacks

Digitalization and Industry 4.0 initiatives require tight integration between complex, heterogeneous and highly complex industrial control systems (ICS) and enterprise information technology (IT). However, the very components that enable digitalization—sensors, connectivity and smart applications—also increase risk.

Digitalization enhances efficiency, improves safety and optimizes production, but it also creates more opportunities for bad actors to penetrate operational technology (OT) environments and to wreak havoc. To secure industrial facilities and ensure safe, reliable production, OT and IT security—traditionally two separate disciplines with different priorities—must merge to share cybersecurity and risk management best practices.

In this article, experts on the front lines of OT cybersecurity risk mitigation share their strategies for making control systems more secure. The firsthand experience collected here comes from experts across a diverse range of industries, including oil and gas, chemicals and refining, and power generation. Their thoughts illustrate the importance of understanding similarities and differences between IT and OT environments. They also share proven experience on adapting IT security controls and best practices to OT environments.

YOU MUST PREPARE TO MEET AN ACCEPTABLE RECOVERY TIME OBJECTIVE

T. Klotz, Covanta

To prepare for a possible targeted attack, the first thing an organization must do is recognize that the threat exists. The more you can educate senior-, executive- and board-level stakeholders, the more successful you will be in responding in those situations. You must get senior stakeholders past the perception of "this will not happen here." Everybody needs to acknowledge the importance of investing to increase the organization's cybersecurity posture. It is critical to balance the financial implications of protection and operational productivity. Organizations must engage in a risk-based discussion to determine how much to invest in security. The investment should align with the risk appetite of the organization based on how much risk the organization is willing to accept.

With stakeholder commitment for stronger security practices, companies must make certain kinds of investments. Segmentation is important to make sure that something does not propagate from one set of operations to another inside the facility and to ensure that it does not propagate from your corporate network into your OT network, or vice versa. You also must manage access so that only the right people have access to the right systems based on their job responsibilities and the parts of the system they must access. Access must be reviewed periodically to ensure that as individuals change positions or leave the company, their access is changed or revoked, as needed.

Having good backups and the right backups becomes the critical factor in a disruptive attack. For example, if a ransomware attack holds a system hostage, how confident are you that you can take the system down and rebuild it to a last known safe state? Have you ever tried to do it? It is important to recognize in the OT space that there are different types of backups. Whether you do differential backups or full daily backups may depend on how frequently information changes and what is needed to restore systems in a reasonable amount of time. In many OT systems, configuration-level information may not change very often, but process control data can change every few seconds.

What is your recovery time objective? How long can you work without the system being available? In an OT environment, that window is typically short—this will be the shortest time that is acceptable to business operations for returning to a fully restored state. Knowing your recovery time objective, you can begin to see how much data you can afford to lose. You can define that for configuration and process control data, and you must determine what those backups look like.

Another consideration in backups is having the data you need to show exactly what happened in the event of an attack. If an attack is the result of a criminal act, you must be able to substantiate where the attack came from and how it penetrated the environment. If cybersecurity insurance is involved, you need forensic information to make your claim. All this needs to be operational data that you capture.

Backup and restoration become part of a disaster recovery plan with a sequence of recovery steps that must be performed. Having that checklist is one thing, but you also must practice it. That is not something you can easily do in a production environment. For critical operations, you might setup separate environment equipment that is identical to the production environment and use that to practice disaster recovery. It can be costly, but your effectiveness at disaster recovery improves with practice. Just as important is the business continuity plan. How will the facility continue to operate as recovery efforts are underway? How long can the facility operate in a manual mode? It is best to have these discussions prior to the event rather than during an event when tensions are high, and the business is suffering.

Responding to incidents in an OT environment is not as simple as in IT. OT

requires a higher level of intrinsic knowledge about facility operations and business impact when there is an incident or disruption to services that support OT. Sometimes, automation can be part of the response plan, but in an OT environment, this is highly situational. Those specific response scenarios must be carefully discussed in the context of specific production operations. If anything is going to be automated, you must know exactly what will happen when that gets triggered. Ultimately, as the lines between IT and OT blur, more people become involved in the response process. Roles and responsibilities must be clearly documented and understood so that response time is not impaired by role confusion. It is imperative that alerts are timely, analysis is swift, business impact is understood, and remediation actions are methodical and planned.

Having a defined incident response plan for your OT systems is imperative, and it must be practiced. IT and OT organizations must work together to protect business operations; there is no time for debate in the heat of the moment. Prepare now so that someone's lack of planning does not become your emergency.

COMMUNICATIONS BETWEEN IT AND OT IS KEY

S. Wilcox, PNM Resources

As a mid-sized utility, we do not have an infinite supply of money and talent to defend against everything a major state actor might theoretically do. What we do have are partners. Cybersecurity, especially for OT, is a team sport. You cannot combat a nation state that has an infinite supply of attackers that they can deploy on a single problem for years. You need a force multiplier, and that force multiplier is partnerships.

A good place to start is with your vendors because they are often your weakest points from an attacker's perspective. You must develop mutual trust with your vendors and relationships with people you can talk to about what is really going on if something happens in your industry. You should also develop relationships with an OT information sharing and analysis center (ISAC). Make friends within your industry.

Furthermore, you must be able to restore systems quickly. If you do not have backups of your systems and your device configurations, you will not be able to quickly recover a process. Instead, you will spend time redeveloping your process controls. You must back up all that information and monitor for changes. This also means looking closely at all your IT-OT integration points.

The integrations are not critical OT systems, but if the integration is not working, the process control will not work. Configurations and integrations must be backed up, and there should be offline backups that are isolated from the rest of the system.

Even with good backups, restoring an OT process can be tricky. For example, many control systems have firmware rather than traditional operating systems. Being able to restore both is important. However, suppose a programmable logic controller (PLC) vendor does a firmware upgrade that turns out to have malware in it. If you install it, you are subject to that problem. You might remediate the device by rolling it to a different firmware. However, if it contains a persistent module, such as a programmable read-only memory (PROM), you may have to replace the entire device. If you have good configuration backups, that is as simple as replacing the device and writing the most recent configuration file.

Another important aspect of OT security is monitoring for configuration changes. You must know when something is changing as soon as it changes. In the OT world, you may not be able to automate a response through security orchestration the way they do on the IT side, but you can have people look at that change if it is unusual.

To make all this work, there must be good, regular communication between IT and OT personnel. If IT and OT people are not talking, you have a much bigger problem than just cybersecurity, primarily because these technologies are converging. They are running on a network that a commodity network switch controls. The networks may be segmented, and in some cases, the OT networks may be separate from enterprise networks. Ultimately, there is a way of getting information into and out of these systems. You must recognize that those points of interconnection are weak points in your defenses. Anything that happens to your enterprise network will over time infiltrate your OT network.

Communication between IT and OT is the key. Try to get everybody together on a semi- regular basis so they can establish friendships and talk the same language. That can be difficult because colleagues live in a world where we have silos of excellence and those silos become isolated. Without good communication between IT and OT, the IT person can see an attack and conclude it did not get in, while the OT engineer can see another failed process control and treat it as just another problem to fix. If people are not talking to each other, they do not realize that the bad guys are attempting to make the process fail.

FOCUS ON RECOVERABILITY

M. Carroll, Georgia Pacific

An essential part of being prepared for a targeted attack is allowing yourself to recover. Many companies recognize the potential risks to the IT side of their business, and the IT-focused tools for detection and recovery are mature vs. OT capabilities. However, companies must recognize just how vulnerable they are—from a business perspective—to OT disruption. If you cannot ship anything, you are not in business.

Being prepared requires being organized to respond quickly and appropriately to whatever happens. That is why we built our holacracy-type organization that provides a blended capability, but people within it know their particular focus. IT focuses on architecture around business systems, business-level communications and cloud interface. Process control is all about process optimization. OT, which we also call manufacturing IT, owns the entire space where process optimization happens, including hardware, software, architecture of the manufacturing space and OT cybersecurity. They all know that their effectiveness together is what makes the product. This awareness means that when something unusual happens, that information cascades through the organization, and everybody who matters is quickly pulled into a conversation. They can decide who needs to focus on what to understand what is going on, really dig into it to understand what it is and then determine what they need to do about it.

The other piece is having the tools to enable recovery. This includes the sophisticated AI-driven tools designed for monitoring and backing up information and configurations in the OT environment. You must have backups in place, running automatically, collecting all the configuration data, fully time-stamped and mul-

tiple copies. When something happens, you can go back and look at that data to determine the exact conditions when it occurred. When it comes time to restore a process, a person in the OT organization will decide what data set is the best one to use for recovery.

The priority in a production process is to restore the process. You have all that data. You use it to determine a recoverable state, wipe out everything and then restore back to where you were before the incident. You isolate data collected after that point, so you do not reinfect yourself. Then, you can set about disentangling what happened. Restore the process before investigating.

It comes down to having two capabilities that become the core of your ability to respond to an attack: the team with all the IT, OT and process optimization knowledge and skills organized in a way that the team can quickly jump in together to respond to a problem; and the tools that continuously back up system data in a way that lets you go back to a good recoverable state. The tools and the people are focused on the same goal: meeting the business objective of restoring production processes as soon as possible.

RECOVERY DEPENDS ON HOLISTIC UNDERSTANDING OF THE OT ENVIRONMENT

M. Chevis, Guffman Ventures

When talking about modern OT control systems and connected OT environments, successfully responding to targeted attacks requires two fundamental things:

1. **Understanding the difference between IT and OT incident response.** You have to recognize that end-to-end incident management and remediation in an OT environment is different than in an IT environment that suffers a malware attack, information theft or unavailability of business systems. You cannot just shut down an OT-managed process without regard for production and safety impact. Restoring basic production capability requires actions tailored to an OT environment rather than an enterprise IT environment.
2. **A good response plan.** You must make sure you have a predefined way to respond to cyber incidents that affect OT systems. Most corporations have an incident response process, but this is not like the IT world where you can back up and easily restore IT systems. OT is different. You must have procedures that you trust to re-create the environment in a state of known integrity so that you can restart your process and your equipment safely.

Tools and capabilities on the IT side can help prevent attacks or minimize the outcomes of an incident that can affect OT systems. You can adapt some of those tools to monitor and analyze OT network traffic to detect potential threats. There are new technologies designed specifically to passively monitor normal OT traffic and send alerts when abnormal communication patterns are detected. One chal-

lenge with these systems is false positives. Humans cannot react properly to floods of spurious alerts. The operator must build trust in the alert part of the system for that to become part of a good incident response process.

Central to any incident response process is visibility into OT assets. If you do not have a trusted picture and understanding of the actual composition of the equipment in the OT environment, effective incident response is challenging. Responding to, stabilizing and recovering from an incident requires a detailed asset inventory. This way, you have high confidence of predictability and assurance that if you take a particular action to remediate equipment or software impairment, you know what the result will be.

Restoring an OT system to a known good state is different from restoring an IT system from a backup. In the OT world, you do not always have mature frameworks for backing up and restoring like you have in IT. Some OT systems are embedded in a way that prevents you from performing image backups. You must rely on configuration backups, yet configurations change. Plants typically have a change management process. It is necessary to find a balance between the efforts you spend backing up configurations of every component in the system and the fidelity of configuration data you need to restore to an operational state. A good OT asset and configuration management solution can help with this.

Both IT and OT must be involved in incident response. Effective incident response requires a blended skillset, but there is a necessary division of roles and responsibilities. The IT side understands network architectures, backup strategies at scale and how to create accessible repositories of backup artifacts. However, when it comes to actual intervention in a process, that can only be done safely by operations.

Intervention on OT systems can be risky if done by pure enterprise IT or an enterprise IT cyber organization because they often do not know what the cascading impact will be of any action they take. With the potential for serious damage to the physical plant and the possibility of increasing the risk profile, actions taken to mitigate and restore must be managed by OT people who understand how the OT systems work, how they interact with the process and why they are designed to work that way.

ONE CHALLENGE IS KNOWING IF SOMETHING IS AN ACTUAL COMPROMISE

J. Laas, PNM Resources

You must do three things to be ready for a targeted attack. These include:

1. **Be aware of threats.** You need a level of threat intelligence specific to your industry and where you operate. An important source of threat information is an ISAC. There are a number of industry-specific ISACs. In oil and gas, we have the ONG-ISAC.
2. **Segregate your systems.** It is important to layer your systems so that you can quickly unplug and isolate islands of functionality. For example, I was involved in a response to the NotPetya attack. When we first learned of this attack, we quickly shut down all PCs and servers, but we had a highly segmented environment that enabled us to do this without disrupting production systems because those systems were fully isolated. This kind of segmentation also helps when restoring to normal operations. Restarting and reconnecting things is often tricky because it is always a big question about when it is safe to restart those systems. With a layered approach, you can do your testing on one segment, and when you are comfortable plugging in that part of the system, you can do that and move on to the next one. That is why it is important to have these separate island modes.
3. **Know what you have.** Maintaining detailed documentation and inventory of your systems are key. This information must be kept updated. System inventory is often part of a crisis management plan. There are technologies that can scan production systems, but you have to be careful with those kinds of tools on control networks because scanning can interfere with the performance of some systems. Segregating your environment can help with this, as well. It enables you to limit access of scanning to those systems that are safe to scan and block access to sensitive control systems.

One challenge in protecting against attacks on OT systems is determining if unusual activity is actually an attack. For instance, you may have a situation where a control system is not operating the way it should, or it is reporting unexpected values on a process. In that case, you should isolate the system, and you may have to call in vendor representatives to have them look at it or install a backup system. For critical systems, you can have test setups that duplicate production systems. This enables you to install a backup and test it. If there is something that should not be there, you must track the change history on that system to see how the change came about and if it was approved. Sometimes, it can take a few days to determine if there has been a compromise.

Collaboration with IT team members is helpful because they can see what is happening in the network through monitoring systems, and they can assist with backups. However, when something needs to be restored on the control system side, only industrial engineers and the vendor can access those systems.

A RECOVERY SOLUTION IS KEY TO OT SECURITY AND AVAILABILITY

V. Ajayi, Ernst & Young

When it comes to attacks on critical infrastructure industries, the stakes are high. These are attractive targets for attackers. Many attacks on these installations involve meticulous planning on the attackers' part and may require collaboration with a malicious insider or a disgruntled employee. To defend against sophisticated attacks, organizations must adopt a multilayered approach to OT network security through strategic implementation of controls across the three domains of people, processes and technology.

For the people strategy, organizations must ensure that anyone who interfaces with the OT network has a baseline awareness of OT cybersecurity. This should include regular refresher training, not just the one-time training for those who work in the OT network. To implement the process strategy, organizations should leverage relevant and applicable industry frameworks and standards to ensure that they have robust processes in

place for every aspect of the OT operation, including inventory management, patch management, events management and incident management. In addition, it is important for organizations to leverage technologies. Many industry leading tools provide real-time information about the landscape of the OT network. The need to have good visibility of all assets and applications domiciled in the OT or process control network cannot be overemphasized. Inventory management is critical. It is imperative to have an account of everything that resides in your OT network.

It is critical to have functional backups and a well-defined disaster recovery process in place. This cannot be overemphasized. It is vital for organizations to have robust backup infrastructure in place, not just onsite, but offsite too, because, in a worst-case scenario, onsite systems and backups can be lost. You must be able to return to operation as soon as possible within your defined recovery time objective. Backup and recovery are key parts of any change management process in an OT network. For example, to install a patch, you should confirm that you have a reliable backup that is updated so that if the patch breaks something, you are able to roll back to a known good state. Having a reliable backup and recovery solution is critical to the security and availability of OT network assets.

Both the IT team and the OT team must work together to achieve a reliable backup and recovery process for the OT network. In fact, they are working toward the same objective, which is to ensure that you can return to production as soon as possible. IT helps provide the connectivity for offline backups, and they play an essential role in saving forensic information and investigating what happened to help prevent those kinds of attacks from happening again. It is vital that the IT team be part of OT incident response planning and testing exercises so they can leverage their understanding of enterprise network dynamics to support any incident that occurs in the OT network. IT should be part of regular drills for incident response so that IT and OT personnel clearly understand their roles and how they should act in case of an incident.

IT and OT teams also must share alert information to fully understand the nature of activities in the OT environment. For example, there can be unusual activity in the IT network that OT should know about. Also, sometimes when you troubleshoot in the OT network, you generate unusual traffic that can raise a flag in the IT network. Open channels of communication should be always maintained between OT and IT teams so that deviations on either side are not normalized.

The importance of this close relationship grows as more OT process control applications move to the cloud. Cloud technology provides many benefits, but it must be adopted with caution in the OT network. To manage risk associated with these new control technologies, testing must be done in a lab environment before deploying anything to production environments. This is another area where IT and OT should work closely together. I see a future where there is no real distinction between IT and OT teams because that is when you can achieve the synergy you need to better protect all the assets in the OT environment.

HAVING A VALID CONFIGURATION BACKUP IS CRITICAL

A. V. Gil-Ortega, Fortinet

The best way to protect against sophisticated targeted attacks is to have strong defenses, but also the ability to recover quickly when an attack happens. To do this, you must have good cooperation between IT and OT teams so you can see activity that is coming from the outside and understand its impact on the OT environment.

For defenses, you need both prevention and detection strategies that reduce the probability and impact of an attack. Prevention includes IT cybersecurity practices designed to keep bad actors out; it also relies on segmentation of the industrial environment. With segmentation, if an attack penetrates, you can isolate it before it affects the whole site. Monitoring and detection are important for recognizing the first signs of an attack on an OT environment. Prevention and detection must work together to meet this kind of threat. With only one or the other, you can easily miss the start of an attack.

If an attack happens, you must identify and isolate it, and then you must restore systems. This requires having data that forensics can use to understand the attack, as well as recent snapshots of the systems you are going to recover. It is important to have valid backups of all operational data and configurations. It is also a good practice to have a separate configuration backup that shows the history of the data and all system changes. Those configurations become the basis for restoring systems and restarting safely.

Without a valid backup, successfully restoring a compromised system is difficult. When backing up OT configurations, you normally run periodic backups and one-time backups whenever there is a significant change. We take a two-step approach that involves backing up system configurations close to the systems so that we can quickly restore them. However, we also have a second backup out of the network in case the entire site is compromised. In this way, once an attack has been mitigated, we can restore proper configurations and restart the systems.

To recover compromised systems after an attack is mitigated, you need access to the last valid backup. This begins by verifying that the source of the backup has not been compromised. You start with local configuration backups and then fall back on the offsite backups. If you cannot verify that the backup is uncompromised, you risk restarting the system in a compromised state. In that case, you must stop all automatic processes and begin manual checks of everything.

To have effective prevention, detection and recovery capabilities, you must have clear knowledge of your control systems inventory. You cannot monitor, detect or restore something if you do not know that it exists. That is only possible with a complete inventory. Additionally, IT and OT must recognize that they need each other to ensure secure operations. IT and OT have different practices, and you cannot pretend they are going to work the same way. In certain areas, they need each other.

For example, given the importance of configuration backups on the OT side, OT engineers can work with IT to make sure there are secure copies in different locations where the OT people can access them. The key is to make backups when significant changes are performed in control systems, and this data cannot be known by IT without proper coordination. Each side must recognize the things that are important to one team are important to the people in the other room, too. HP

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Resourcefulness solves problems

In the now almost 32 yr since this monthly column was first published, we have had more than a few occasions to highlight the pitfalls of trial-and-error solutions in plants that process toxic, flammable or explosive materials. So, why not mention two fine examples where resourceful action by subject matter experts (SMEs) has solved problems?

A bumper sticker becomes a coupling stretch monitor. Forty-seven years ago, in late 1975, an SME with a well-balanced combination background that was both practical and theoretical participated in commissioning an important centrifugal compressor train at a facility in France. He knew that the long span between the respective shaft ends of the large centrifugal compressor and its steam turbine driver was bridged by a diaphragm coupling. In modern turbomachines, the taper-bored hubs of diaphragm couplings—virtually identical to **FIG. 1**—engage the tapered shaft ends with a strong interference fit, and keys are no longer used. Although shown with a keyed fit to shafts identified as #28 and #30 on the patent drawing reproduced in **FIG. 1,** the machinery train in France was one of the first that dispensed with keys. The two coupling hub bores were dilated by hydraulic means and forced up on their respective shaft ends until an interference fit of 1.5 mm/m–2 mm/m (0.0015 in./in.–0.002 in./in.) was reached.

However, compressor shafts tend to grow when they operate surrounded by hot process gases. Steam turbine shafts will grow even more because steam is usually extremely hot. At standstill, and before gases and vapors are admitted to driver and driven machines, both shafts are at stable ambient conditions. The distance between shaft ends (DBSE) will be at its maximum and the two diaphragms in **FIG. 1** will be "stretched" a small amount. Once the compressor is brought up to speed and is performing its compression duty, the shafts of both driver and driven machine will have thermally grown, and the DBSE will be at its minimum. Hopefully, and under stable operating conditions, the two diaphragms will be "normal and relaxed," meaning neither stretched nor pushed towards each other.

But how did the SME verify that "normal and relaxed" had been achieved as pre-calculated and anticipated? He took two shiny automotive bumper stickers—auto insurance advertisement stickers—(2 decals, items "24" in **FIG. 1**) and with a fine-point felt-tip marker drew progressive steps "a/b/c/d" 0.040 in. (1.0 mm) apart. With a pen knife, he cut these steps before bonding each decal to the coupling spacer right under the guard plates "20." At "22," each of the two guard plates has a radial clearance of 0.5 mm (0.020 in.) relative to spacer tube "12." At normal speed, observation with a stroboscopic light showed how many steps were covered (or uncovered) in the plus or minus directions. The coupling diaphragms were doing fine with the machine at stable load and temperature.

How remembering Archimedes solved a leakage problem. During commissioning of several pusher centrifuges (**FIG. 2**), it became immediately evident that there was excessive and intolerable leakage from the annular opening between the stationary inlet elbow and the rotating centrifuge basket. The courteous Japanese startup managers asked if the normally U.S.-based SME (who had visited the Swiss centrifuge manufacturer during the mechanical run test phase) would please design a mechanical seal that would solve the problem.

Designing and fabricating a special mechanical seal would take weeks, the SME thought, but declining the request would have been a disappointing refusal. The SME began by sketching a mechanical seal with parts made of bronze, polytetrafluoroethylene (PTFE) and stainless-steel springs. The refinery's machine shop fabricated the components overnight, and a machinist crew was asked to install the parts in time for a pre-noon restart attempt on the following day. Unfortunately, the

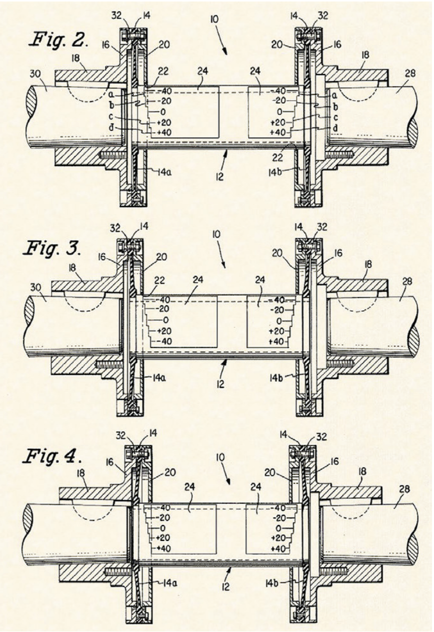

FIG. 1. A page from a long-expired coupling decal patent.

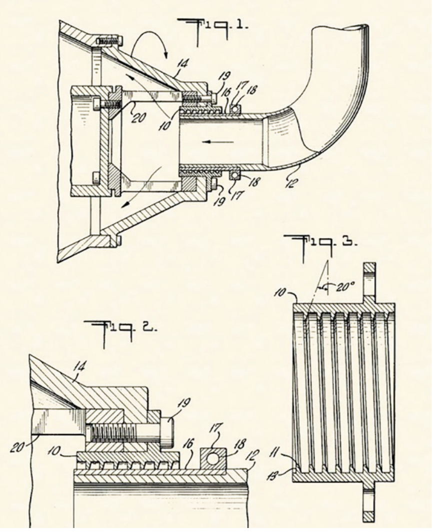

FIG. 2. While an Archimedes screw has the helix on the outside, this "Archimedes bushing" has the helical screw flight on the inside.

SME's mechanical seal was a dismal failure. It exhibited unexpected outward pumping action; the leakage rate of noxious paraxylene was now worse than before.

However, new lessons had been learned on the spot, and no effort was made to tweak yesterday's obviously inferior design idea. In a true brainstorming session attended by the SME, local engineers and technicians, the team members nudged each other towards thinking "out-of-the-box." An internal, square-threaded, helical-flight bushing (**FIG. 2**) took its inspiration from Archimedes and was sketched, machined and tested within 24 hr, completely solving the problem of product leakage. A patent was obtained.

In short: There is a time and a place for innovative in-house thinking, especially when time is of the essence. However, at no time should this thinking disregard safety. Also, wherever possible, reliability professionals will benefit from involving other potential contributors in the deliberative processes. Finally, we should also remember that tweaking inferior ideas is not problem-solving and will rarely substitute for science-based designs. Archimedes led the way in 200 B.C.E.

Identifying potentially resourceful employees. Not every employee will be self-motivated or interested in becoming a resourceful problem solver, and an employer can live with that. Still, even during a job interview a graduating engineer would be wise to explore his or her projected role and consider details. Soon after starting work, the engineer should ask the manager for a written role statement. If no such statement is forthcoming, the engineer may put his or her understanding on paper and ask the responsible manager for review, input or concurrence. Unless agreement on the engineer's role has been reached, "performance exceeding expectation" is rather improbable.

Likewise, during the job interview, an engineer about to graduate should ask about the training opportunities made available or endorsed by the prospective employer. The interviewee's goal must involve professional growth and learning. Learning is obviously a two-component process. While one party offers it and the other absorbs it, the ultimate benefits are shared by both.

As an example, a company could identify a self-motivated employee and ask this employee if he or she would be willing to be the custodian of an electronically stored and searchable engineering library dealing with turbomachinery, pumps, gears, shaft couplings, etc. During periodic performance appraisals and reviews, his or her progress towards becoming a solid contributor could be examined. HP

HEINZ P. BLOCH resides in Montgomery, Texas. His professional career commenced in 1962 and included long-term assignments as Exxon Chemical's Regional Machinery Specialist for the United States. He has authored or co-written more than 780 publications, among them 24 comprehensive books on practical machinery management, failure analysis, failure avoidance, compressors, steam turbines, pumps, and the just released "Optimized Lubrication, Oil Mist Technology, and Standstill Protection," (DeGruyter, Berlin/Germany, ISBN 978-3-11-074934-2). He holds BS and MS degrees (cum laude) in mechanical engineering and is an ASME Life Fellow with life-time registration as a Professional Engineer in New Jersey.

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Case 115: A method for analyzing catastrophic type failures

A portion of my consulting career involved investigating catastrophic type failures on machines, pressure vessels and structures in the industry. Causes and solutions were required quickly so the equipment could restart safely, and production resumed.

While collecting and analyzing the damaged and mangled parts, my involvement has been to propose and test theories on the most probable causes. Some form of analytical analysis is used based on the available data, and I share this information with the investigating team and modify this as I obtain new information. This helps focus the team on all the confusing debris and input from interviews and increases understanding of what might have happened. Testing the theory in this way will either validate it or provide reasons to dismiss it. This is similar to what is done using the scientific method. In the scientific method, a theory, hypothesis or best guess on how something has occurred is proposed based on the data available. As additional data are gathered, such as interviews/metallurgical/material evidence, they are tested to see if they agree with the theory.

It has been my experience that several events usually have to occur to result in a failure–lack of or delayed maintenance, overloads due to exceeding operating or design specifications, faulty repairs or modifications are typical causes. Major design deficiencies from the manufacturer are rare if built to their specifications. A long history of successful operation before a failure suggests this.

Consider theorizing the wreck of a 2,000-hp gas engine-reciprocating compressor as shown on **FIG. 1.**

The following theory was proposed after reviewing the data shown in **TABLE 1,** which is only a small sample.

The slipper bolts had become loose over time, which was the ticking heard by the operator. The dynamic load impacted the body of the bolts and caused them to fail in shear. The loose slipper wedged itself against the piston rod, and when the crosshead came forward, it sheared the rod. This was the bang the operator heard.

Others suggested several other causes; however, the data did not support them. The solution was to install dowels to secure the slippers in addition to the bolts. This eliminated the cyclic load on the bolts, which loosened them. Periodic torque inspections were recommended as a maintenance item. No such failure occurred for 15 yr after the modification, at which time the machine was replaced with a centrifugal compressor.

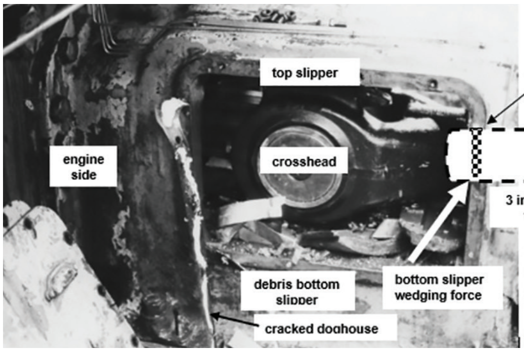

FIG. 1. Wreck of doghouse section

NOTE

Case 114 was published in *HP* October 2021. For past cases, please visit www.HydrocarbonProcessing.com.

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TONY SOFRONAS, D. Eng, was the worldwide lead mechanical engineer for ExxonMobil Chemicals before retiring. He now owns Engineered Products, which provides consulting and engineering seminars on machinery and pressure vessels. Dr. Sofronas has authored several engineering books and numerous technical articles on analytical methods.

TABLE 1. Data to help verify theory

Observation	Source
1. An experienced operator heard an unusual continuous ticking and then a bang just before the failure.	Interview.
2. The compressor piston rod had sheared off due to wedging of the broken slipper.	The calculation was performed along with metallurgical analysis—high wedging force available to do this.
3. The slipper bolts had all sheared off due to dynamic impacting.	The calculation was performed along with metallurgical analysis. Loose hold-down bolt torque and shear probable.
4. The slipper bolts had not been torque checked for many years.	Maintenance records and interviews.
5. There were two bolts with low torque on the six other slippers out of 24 bolts.	Torque wench.
6. The safety wire was still in place.	Observation.
7. The slipper was cast iron and broken into pieces, and the 3-in. diameter piston rod had sheared. The nut was tight.	Observation and torque wrench.
8. This type of failure had happened to the same type of unit at another division.	Interview with a colleague at a gas compression plant.

HELION SARDINA, CHIEF COMMERCIAL OFFICER
Lummus Technology

What is the future for the HPI?

Hydrocarbon Processing sat down with Helion Sardina (HS), Chief Commercial Officer, Lummus Technology. In this interview, Mr. Sardia offered his market insight on the COVID-19 pandemic, the process technology industry now and in the future, and its role in the energy transition.

How did COVID-19 affect the hydrocarbon processing industry in 2021?

HS: When COVID hit in 2020, it plunged the refining and petrochemicals industries into a deep downturn, except for products such as polymers that are tied to the medical sector. Fortunately, 2021 was not as severe as 2020, and the industry is close to a worldwide recovery.

The pandemic resulted in lower plant utilization rates and caused some construction projects to be delayed. Due to our company's diversification and longer-term planning and execution windows for customers, we fared much better than most other players in our industry. We did, however, experience challenges and delays in being able to deliver proprietary equipment for ongoing projects and reloading catalysts for our licensees.

Do you foresee similar impacts in 2022?

HS: Based on our current pipeline, we are seeing previously delayed projects now moving forward. If efforts to curb the pandemic are successful, it will bring our global markets back into balance. As a result, we anticipate increased consumer activity, which fuels growth in our business and industry.

What are some of the key drivers and technology developments shaping the market today? In the next 5 yr?

HS: We need to transform current technologies and develop new ones that minimize environmental impact and contribute to the circular economy. It is also important that we provide refiners solutions that help them pivot away from declining transportation fuels by repurposing existing assets to meet the growing demand for petrochemicals.

Lummus' efforts include developing and administering licenses focusing on world-scale crude-to-chemicals projects. We have added energy conservation components to our portfolio, such as gas turbines in our ethylene crackers. We have teamed up with New Hope Energy, which has a leading plastic waste conversion technology that produces pyoil that can be co-fed into a refinery or petrochemical plant with fossil fuel-derived feedstocks. With this partnership, we have applied our expertise to scale up the plant design to ensure it can have a significant impact on the end-of-life plastics challenge.

What role will process technologies and companies like Lummus play in the energy transition?

HS: Customers across all regions are requiring more environmental, social and governance compliance, and a lower carbon footprint at their existing facilities.

In late 2020, Lummus established Green Circle to focus on the circular economy and provide energy transition solutions. We have developed new technologies related to the energy transition, while continuing to enhance existing technologies that produce renewable and cleaner fuels, lower the carbon footprint of our customers' downstream investments and help facilities co-process circular products.

Our industry has a long and lasting legacy of making things work. We know how to convert hydrocarbons into fuels and consumer products. This is one thing that has not changed. Therefore, process technologies and the companies that license them can build on this legacy in the energy transition.

How does the process technology industry need to evolve going forward?

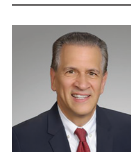
HS: Our licensees are under pressure to reduce emissions and become more environmentally friendly, and we can help them lower the carbon intensity of their plants.

We must provide every one of our licensees the most efficient plant design, which translates to the lowest impact to the environment. This holistic approach can reduce the overall carbon footprint while also improving plant operating expenses. We need to continue minimizing feedstock consumption, optimizing energy use, extending the use of catalysts and reducing emissions.

On the surface, those do not typically sound like they go together—plant operations and environmental cleanliness. However, process technology licensors have historically produced technologies that have helped customers increase efficiency, emissions and waste reductions and savings in their operations.

Our industry can also evolve by accelerating digitalization, an area where we have been slower than some other industries in adopting. At Lummus, we formed a joint venture with TCG Digital. The JV works with our existing customers and prospects to implement digital solutions for their refining, petrochemical and gas processing assets, as well as across the hydrocarbon processing value chain.

We see opportunities to further the digitization of our industry—from performance monitoring, system optimization and remote management of facilities, among others—to empower industry leaders with smarter ways to do business. HP



HELION SARDINA has spent his entire career with Lummus Technology, serving in different management roles of increasing responsibility. At present, Mr. Sardina is Lummus' Chief Commercial Officer, where he is responsible for overseeing all aspects of sales and business development for the company.

B. BURTON, Emerson Automation Solutions, Shakopee, Minnesota

Reduce maintenance costs by optimizing pressure transmitter calibration intervals

Pressure instrumentation is crucial for the functionality and safety of hydrocarbon processing facilities worldwide. Many of these facilities deploy hundreds or even thousands of pressure devices to accurately monitor and maintain optimal pressure levels in various applications. With so many of these devices in one plant, it is essential to minimize servicing and maintenance costs. Optimizing the calibration frequency of pressure transmitters is one way to achieve significant cost savings, helping producers maintain affordability and allowing them to achieve greater operating returns. Unfortunately, legacy best practices and broadly applied internal standards have created calibration maintenance schedules with shorter than required intervals, resulting in high and unnecessary maintenance costs.

Shorter maintenance windows generate high costs for producers in the form of extra time, money and resources spent, as well as additional risks to personnel as technicians spend more time in the field. When operators can estimate more accurate calibration frequencies for pressure transmitters, these maintenance costs may be considerably reduced. If operators take another look at their calibration schedules, they will find that modern pressure transmitters do not necessarily demand such short calibration intervals. It is possible to estimate a more precise calibration interval by performing a few simple calculations.

Pressure transmitter technology has evolved dramatically in the last two decades. Advanced instrumentation enables operators to depend on these devices for longer intervals without sacrificing accuracy or reliability. Although these intervals are not commonly stated in a pressure transmitter's literature, there are ways to extrapolate them from published specifications and data so that operators can safely maximize the time between maintenance cycles.

$$\text{Total Probable Error} = \sqrt{\underbrace{\underset{\text{Reference Accuracy}}{\left(X_r\right)^2}}_{\text{Accuracy at Calibrated Range}} + \underbrace{\underset{\text{Ambient Temperature Effect}}{\left(X_t\right)^2} + \underset{\text{Line Pressure Effect}}{\left(X_p\right)^2}}_{\text{Installation Effects on Accuracy}}}$$

FIG. 1. Root sum squared tolerance method used to calculate TPE.

The driving factors and costs of frequent calibration. The processing industry's most critical applications rely on pressure technology. Pressure instrumentation devices can also be platforms for level and flow. They must be accurate and reliable so it is easy for operators to assume that calibrating them more often will achieve better performance. However, current best-in-class transmitter technology does not require such short calibration intervals, and frequent maintenance may do more harm than good. In addition, the manual routine of calibrating each transmitter uses considerable resources and disrupts the processes that depend on them. Devices are located everywhere around a plant, in hard-to-reach places, making the work all that much more difficult. Technicians must be in the field for more extended periods, shutting down transmitters to attach a calibration device and checking if the reading is accurate.

Older devices were not as stable as the sophisticated equipment available today. Many operators use legacy best practices and standards carried over from older technology to select the calibration frequency on their new technology. Applying broad regulatory requirements or ambiguous performance parameters written to cover field devices that have poorer capabilities can also contribute to operators overscheduling maintenance on advanced pressure sensing technology.

How to calculate accurate calibration intervals. Manufacturers do not usually recommend calibration intervals for their products, but published specifications and application-specific data can be used to determine a reasonable estimate. There is a five-step process to establish the correct calibration interval for a pressure transmitter.

Step 1. Step 1 involves determining the performance required. What are the general monitoring, reporting, recordkeeping and verification requirements for the pressure transmitter? Typically, allowable uncertainties as a percentage of calibrated span are a function of the criticality of the application with safety and plant efficiency at 0.5%, regulatory control at 1%, supervisory control at 1.5%, and monitoring and optimization at 2%.

Step 2. This step defines the operating conditions. Changing conditions impact performance. Input the temperature range and variations the device will be exposed to in ambient conditions. For a differential pressure measurement, it is important to input the line pressure variable.

Step 3. Step 3 is a total probable error (TPE) calculation utilizing a root mean squared method (FIG. 1). The square of the reference accuracy is added to the squared temperature effect and added to the squared span static pressure effect. The square root is then taken to determine the TPE. For a static pressure device measuring absolute or gauge pressure, the pressure effect is zero.

Step 4. This step determines the stability of the output. The stability specification is a key variable and the final input to the calibration frequency calculation. Stability specifications can vary widely depending on the performance class of the instrumentation. Industry-leading pressure transmitters have stability specifications of 10 yr–15 yr. The higher the

stability specification, the longer the calibration interval will be.

Step 5. Step 5 takes all these factors into account for the final calibration frequency calculation (Eq. 1).

$$\text{Calibration frequency} = (\text{Required performace} - \text{TPE}) / \text{Stability per month} \quad (1)$$

Required performance minus the TPE is divided by stability per month to determine the proper calibration interval. Be sure to keep units of measure consistent across the equation. The two variables that affect the outcome most are TPE and the device's stability specification. These factors make a significant difference in the expected calibration interval.

A real-world example of significant maintenance cost reduction. By utilizing these equations, operators can obtain and analyze interval data at a more granular level. These precise calculations may result in meaningful savings and maintenance cost reductions for facilities. This method has already proven beneficial to producers. One such real-world example impacted a major U.S. refinery.

This refinery was unsure how to perform calibration in a way that would satisfy the U.S. Environmental Protection Agency's new greenhouse gas requirements (Title 40 CFR Part 98—Mandatroy Greenhouse Gas Reporting). They decided to take the most conservative approach to their calibration program, performing expensive annual calibrations to no benefit. Often, technicians would apply pressure and find devices well within their performance limitations.

The refinery had installed cutting-edge pressure transmitter technology but was not taking advantage of its benefits. With a stability specification of 15 yr, their transmitters could remain accurate and reliable for much longer periods of time. Extensive field data was gathered to test this theory. Every time a technician calibrated a device, they noted and tracked the data.

A database was built to analyze the pressure transmitter calibration data. They found that out of 89 devices, only one required recalibration during the 6-yr data collection period. Every other device was within specifications. The device that fell outside of accuracy requirements was a much older, legacy technology. The refinery decided to replace that device, along with several other older transmitters, with the newer model.

Based on this data, the refinery moved 89 differential pressure transmitters from a 1-yr calibration interval to a 3-yr calibration interval, while remaining greenhouse gas compliant. Not only did this change free up resources and keep personnel from doing unnecessary testing in the field, but they also saved approximately \$22,000/yr. Ongoing field data suggests the refinery may extend the interval to 5 yr. Refinery personnel are making considerations to extend this program across their installed base of more than 4,000 devices.

High-performance pressure transmitters deliver longer calibration intervals. Facilities may want to embark on their own field studies to test these estimated calibration intervals and further justify the move towards optimization of calibration frequency plantwide. The activity is worthwhile, with potential cost savings of tens of thousands of dollars each year.

The extended stability specifications available in advanced pressure sensing technology heighten the need to reevaluate traditional calibration schedules. There is no need to spend resources on constant physical checks when instrumentation can withstand production environments, while remaining within specifications for much longer windows of time. These maintenance cost savings may even offset the initial purchase price of newer models.

As measuring equipment becomes more robust and dependable, operators must reconsider outdated methodologies that drive unnecessary expenditures. Analyzing current calibration procedures for pressure instruments throughout a facility is a step in the right direction and could save thousands of dollars a year in operating costs. HP

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Optimize refining operations using plant digital twin based on molecular modeling

A new era of the energy revolution has arrived with the rapid growth of innovative technologies that utilize alternative energy resources, such as electricity, solar, wind and hydrogen, and an increasing demand for sustainable use of natural resources.

Refiners are facing unprecedented challenges in maintaining their profit margins. For example, the demand for fuels, especially gasoline, is decreasing significantly due to the widespread adoption of new vehicles that run on various fuels. In the next decade, the refining industry is expected to shift from a fuel-oriented industry to a raw material-oriented industry; therefore, crude-to-chemicals is one of the major technology paths for refining operations. Moreover, strict governmental and societal regulations on carbon dioxide (CO_2), methane (CH_4), sulfur oxides (SO_x) and nitrogen oxides (NO_x) emissions are pushing refiners to reduce carbon emissions and upgrade to alternative feedstocks (e.g., biomass, pyrolysis oils) together with petroleum fractions. **FIG. 1** shows a typical refining flowsheet for crude-to-chemicals.

This article will describe how molecule-based process simulation can help address the challenges of achieving crude-to-chemicals process benefits, developing refining processes to process alternative feedstocks, such as biofuels, and more.

Process description. As shown in **FIG. 1,** crude oil and biofuels both feed into a refinery. Crude oil first passes through a crude distillation unit (CDU), where it gets separated into various boiling fractions: straight-run naphtha, distillate, gasoil and atmospheric residue (AR). Biofuels are upgraded via a hydro-deoxygenation (HDO) unit and converted to paraffinic distillate materials. The effluent of the HDO is mixed with the petroleum distillate from the CDU and upgraded via a hydrotreater unit to produce high-quality diesel. Alternatively, the product of the HDO can be combined with the distillate and/or gasoil from the CDU and cracked through reactors (e.g., a hydrocracker or HCR) to produce HCR naphtha. Moreover, naphthas from different plants (e.g., straight-run naphtha and HCR naphtha) can be further upgraded through reactors (e.g., reformer, steam cracker) to produce more desirable products.

Challenges with the crude-to-chemicals process. Unlike the traditional refinery, the products of the crude-to-chemicals refinery are not solely fuels (e.g., gasoline, diesel, jet fuel)—the naphtha fractions are the feedstocks to petrochemical plants (e.g., ethane cracker and aromatics production). The typical scope of traditional refinery optimization is a single unit (e.g., CDU, reformer, hydrocracker) or several units. For the crude-to-chemical scope, the operator must move beyond the unit level optimization to consider optimization across both the refinery and the chemical plant.

However, the traditional lumped model approach cannot solve the optimization of the process simulation flowsheet shown in **FIG. 1.** The lumped models used in refinery process simulations (**TABLE 1**) usually define their species by physical properties, such as boiling point, specific gravity or solubility; however the species used to model a petrochemical plant are molecular components.

For example, a detailed paraffin-isoparaffin-olefin-naphthenes-aromatics (PIONA) carbon number breakdown is required to model ethylene cracking and/or aromatic chemicals. The lumped model cannot provide functions to precisely propagate the molecules from the refining units to the chemical units. Furthermore, the new feedstocks like the biofuel feeds shown in **FIG. 1** are beyond the definition of the lumped model and cannot be described using the existing lumps. It is also necessary to describe refining upgrading processes at a highly granular level to get a better understanding of the chemistries, and provide optimal operations that comply with regulations on carbon and other emissions. This motivated the development of a novel approach to address this issue: molecular level modeling.

Molecular level modeling. Because molecules are the fundamental elements in any refining and chemical process—and can reveal the nature of chemical conversions—this novel molecular modeling solution approach provides refiners an optimal solution to address the aforementioned challenges. It begins with the company's molecular characterization (MC) technology,[1,2] which has established a molecular library con-

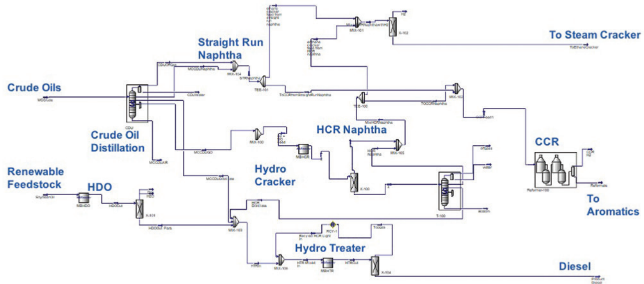

FIG. 1. Representative refinery models for crude-to-chemicals.

TABLE 1. Benefits of MB reactor

Customer interests	Required specs	Conventional reactor	MB reactor
High-accuracy model of selected properties; integration of petrochemical	Identify branched isomers of PIONA	Only supports empirical correlation	Includes detailed branched molecular species
Deep understanding and optimization of refining chemistries	Describe more realistic kinetic and mechanism	Not supported	Detailed reaction path: isomerization, HDS
Catalyst design and development	Intrinsic kinetic parameters	Not supported	Obtains more intrinsic kinetic parameters of catalyst; less dependent on flowsheet, feedstocks
Resid processing	Describe heavy-end conversion	Not well-supported	Includes archipelago resid structure and reactions
Apply user's in-house kinetics and reactor models	Allow users to add or edit components, reactions, kinetics	Not supported	Automates code generation for user's in-house components, reactions, kinetics
Propagation of detailed high-quality data of reactors to proprietary planning software[c]	Estimate molecule-based structural properties	Not supported	Supports estimations of molecule-based structural properties
An optimal way to provide high-quality data to machine-learning model	Process data (measurable and non-measurable) of all necessary dimensions	Not well-supported	Inherently provides densified data and reveals most important data among complex chemistries

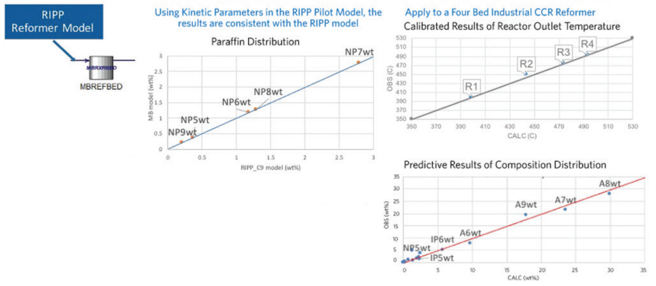

FIG. 2. Customized MB reformer model in MB reactor.

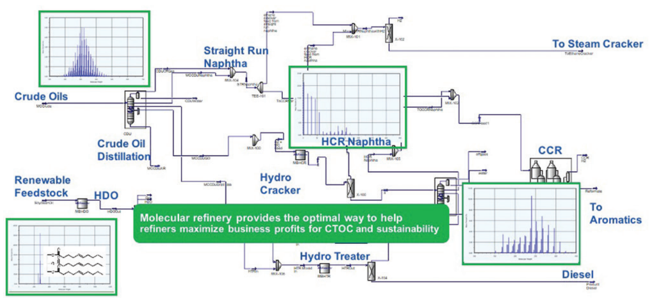

FIG. 3. Molecular sustainable HPR across a wide range of refining and chemical models.

sisting of 22 distinct molecular compound classes and more than 60,000 molecules with a comprehensive rigorous thermodynamic property package driven by proprietary physical properties models[a]. By leveraging the information from that library and the analytical data of crude oils, MC can characterize the molecular compositions of crude assays so that they optimally match molecular and bulk measurements.

Using the molecular assays from MC as the initial values, a general molecule-based (MB) reactor[1,2] modeling framework was developed to simulate and optimize molecular reactor models and propagate the molecular information throughout the flowsheet. The MB reactor can organize O (10,000) molecular components in refining streams as a set of homologous series and describe O (10,000) distinct reactions in a given refining or petrochemical reactor. Employing a linear-free energy relationship (LFER)[3,4,5] technique, refiners can simulate and calibrate a complex refining reactor model with only O (30) kinetic parameters and obtain the high-fidelity and high-granularity molecular data of the products. The benefits of the MB reactor are summarized in **TABLE 1.**

An MB model builder was developed for users to include their own detailed molecular models into the MB reactor without requiring hard coding. The MB model builder has been used to develop a MB hydrocracker/hydrotreater model in terms of more than 2,400 molecules and 5,700 reactions.[6] Refiners can use this unit operation to simulate and calibrate a hydrocracker model and propagate high-fidelity molecular data through a flowsheet of a refinery.

SINOPEC Research Institute of Petroleum Processing (RIPP) leveraged its own molecular reformer models to the MB reactor via an MB builder and successfully simulated a four-bed industrial CCR plant, as shown in **FIG. 2.**

In addition to modeling petroleum feedstocks at the molecular level, the MB reactor can also model alternative hydrocarbon feedstocks such as biomass, lignin, cellulose, hemicellulose, plastics, coal, etc., allowing the MB reactor to support modeling sustainable feedstocks in a refinery flowsheet. A green diesel[7] model to upgrade biofuels was developed via an MB reactor.

Takeaways. Molecular modeling can address the upcoming challenges for refiners, and it is possible to track the molecular compositions of any stream in a flowsheet across a wide range of refining and chemical models, as shown in **FIG. 3.** Users can utilize high-fidelity molecular data, not only for a sustainable crude-to-chemicals steady-state simulation or calibration, but also for the integration with online optimization or for planning and scheduling.[8,9] Molecular modeling provides an optimal plant digital twin solution to optimize refining operations for decades to come. HP

NOTES

[a] AspenTech PC SAFT models
[b] Aspen HYSYS Petroleum Refining V12.0
[c] Aspen PIMS

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Web-based software for predicting crude compatibility and optimization for increasing heavy oil processing

Refineries in oil-importing nations typically process a blend of crude oils, rather than a single crude oil, to ensure that an optimum product mix can be obtained at the minimum costs. To increase margins, refineries are looking for ways to co-process heavy crude oils with light crude oils.[1–4] Heavy crude oils contain high amounts of paraffins or asphaltenes. High paraffin content results in high viscosity and high pour point, making transportation difficult. Conversely, high asphaltene content causes precipitation, flocculation, instability and incompatibility challenges during processing. This severely affects process equipment like heat exchangers, pumps and tanks.[5–8]

The current benchmark process to determine compatibility of blending two crude oils consists of 9–10 standard laboratory-based test methods, out of which 3–4 are required to be done as per their applicability range. These tests can take weeks to complete. At present, there is no standard practice to check the compatibility parameter in advance; rather, chemical dosing is done to prevent incompatibility-related problems in refineries. Therefore, increasing co-processing of heavy oil components is a real challenge.

To increase the heavy oil content in the mix of crude oils, and for suitable oil selection for co-processing, refiners encounter several common problems on day-to-day operations. These include:

- Incompatibility/stability issues when the crude oils are blended
- High viscosity of the blend
- High pour point of the blend
- High sulfur content in the blend
- High acidity and nitrogen content
- Low distillate yields and availability for feedstock choices.

The objective of this article is to provide a quick and effective method for predicting crude oil blend compatibility, as well as for optimizing heavy oil processing, using a prediction model software[a]. The prediction model is based on the measurement of a few bulk physical parameters, which are conventionally and regularly analyzed in a refinery's quality-control laboratory. This analysis typically takes less than 1 hr, with no additional tests required, which enables refiners to quickly make blending decisions.

In contrast to conventional methods, the present subject matter does not require comprehensive laboratory testing for compatibility and blending, which otherwise normally takes several weeks. Using the prediction model software[a], operators can increase heavy oil processing and substantially eliminate operational problems related to asphaltene precipitation caused by crude blend incompatibility. Furthermore, conventional laboratory test methods can optimize the blending of only two crude oils at a time. To optimize a blend of three crude oils, a compatible blend test of two crude oils must be obtained, followed by a compatibility test of the first two blended crude oils with the third. If additional crude oils need to be blended, then the compatibility checking and blend optimization becomes even more complicated.[2]

Therefore, another objective of this article is to devise a methodology for compatibility prediction and for the optimization of blends having any number of crude oils. The focus of this work can be used for increasing heavy oil processing and can help eliminate problems caused by crude oil incompatibility.

Methodology. Asphaltene precipitation has been a common problem in refineries. It occurs due to incompatibility of the crude mix, especially when the heavy oil fraction increases in the blend. The quick and reliable prediction of crude oil blending compatibility is critical for the best selection of crude oil blends.

The compatibility of crude oil blends can be estimated using the following tests: the colloidal instability index (CII), the colloidal stability index (CSI), the Stankiewicz plot (SP), qualitative-quantitative analysis (QQA), the stability cross plot (SCP), the Heithaus parameter (or parameter P), heptane dilution (HD)/toluene equivalence (TE), the spot test and the oil compatibility model (OCM), among others.[9–13] All these experimental methods are based on the physical model of asphaltenes and their solubility with

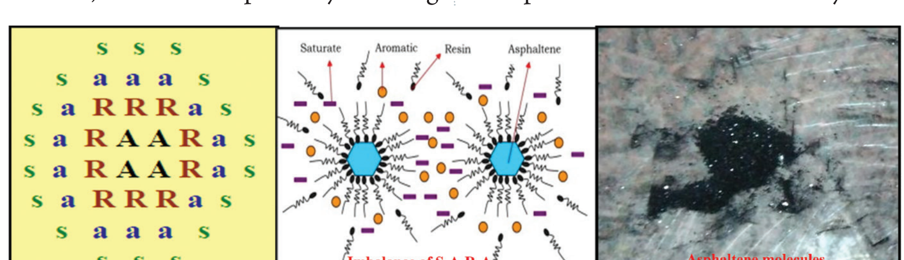

FIG. 1. Physical model of asphaltene precipitation.

TABLE 1. Comparison of the K model vs. different compatibility tests: Unstable (U), stable (S) or metastable (M)

Crude oils	CII	CSI	SI	QQA	SP	SCP	Spot test	Composite result	*K* model results	Refinery experience
AH	U	U	U	U	U	U	U	U	U	U
AEL	U	U	S	M	S	S	S	S	S	S
AL	M	S	M	U	S	S	S	S	S	S
Ratawi	M	U	U	U	U	U	U	U	U	–
SB	U	U	S	S	S	S	S	S	S	S
RG	S	U	U	U	U	U	U	U	U	–
Murban	M	S	S	M	S	S	S	S	S	S
RJ	U	U	S	U	U	S	U	U	U	–
ML	M	S	S	M	S	S	S	S	S	S
Erha	U	S	S	S	S	S	S	S	S	S
MH	M	S	S	S	S	S	U	S	S	S
Antan	M	S	S	M	S	S	S	S	S	S
Mascila	M	S	S	U	S	S	U	S	U	S
Kuwaity	M	M	U	U	U	S	S	M	S	S
AM	M	S	M	U	U	S	S	S	S	S
NB	M	S	S	M	S	S	U	S	U	–

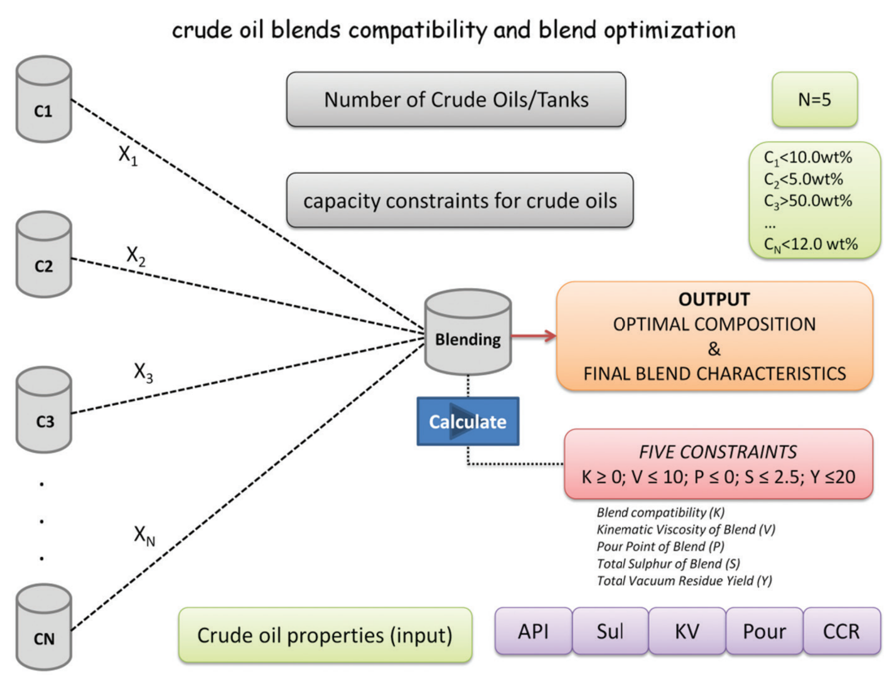

FIG. 2. Optimization module in the prediction model software[a].

other components present in the oil system. The typical physical model has been depicted in **FIG. 1.** These experiments can take a long time to conduct—1 wk per blend of two crude oils at a time—and are, therefore, tedious.

Proprietary prediction model software[a]. A novel tool for predicting the compatibility of crude oil blends and blend optimization for increasing heavy crude oil processing has been developed. This model uses the four physical parameters of crude oils—sulfur, carbon residue, API and kinematic viscosity—as input for blending optimization.[14–23]

The model is developed by coefficients obtained via regression analysis between the ratios of the physical parameters of known crude oils, and by the composite compatibility measures determined from multiple compatibility test results of the known crude oils. The model's equation is provided in Eq. 1. This effectively correlates characteristics of asphaltene molecules and their behavior. The content of aliphatic carbon attached to the aromatic core of asphaltene in a heavy crude oil is the primary deciding factor for determining crude blend compatibility when blending with light crude oil. Higher aliphatic carbon attached to the aromatic core of asphaltene in heavy crude imparts instability when blending with light crude oils.

$$\begin{aligned} K = {} & k1 \times (1/C) + k2 \times (C/A) + \\ & k3 \times (C/S) + k4 \times (\log(S)/C) + \\ & k5 \times (A/S) + k6 \times (S/V) + \\ & k7 \times (V/A) + k8 \times (V/C) \end{aligned} \quad (1)$$

where $k1$–$k8$ are the coefficients of regression. When $K \geq 0$, the crude oils are compatible; when $K < 0$, they are incompatible.

The proprietary prediction software[a] accurately predicts the composite results of all comprehensive laboratory test methods within a few minutes. The prediction software enables refiners to predict compatibility of multiple crude oils (up to 10) within a short amount of time.

The parameters for blend optimization are blend compatibility, blend viscosity,[24] blend pour,[25] blend acidity,[19] blend sulfur, blend nitrogen[19] and total distillate yield.[19] The optimization module also considers crude oils/tank storage availability (**FIG. 2**).

Validations. Seventy different crude oils have been used for the development of the model prediction software, while 16 neat crude oils and 14 crude oil blends have been used for validation (**TABLES 1–2**). The *K* value was determined based on the compatibility model in Eq. 1 for 16 crude oils. The results of the compatibility parameter *K* were compared with the compatibility results based on saturate, aromatic, resin and asphaltene (SARA) analysis and spot tests.

It is known that any single test method is inadequate to make accurate decisions regarding the compatibility of crude oils and blends. In this case, individual and composite results of all known laboratory test methods have been considered to validate the *K* model. In addition, some of the 16 crude oils were also processed

TABLE 2. Validation of the K model vs. different methods for blends: Unstable (U), stable (S) or metastable (M)

Crude oil blends			Compatibility											
SN	Heavy oils	Light oils	*K*	Compatible blends	CII	CSI	SI	QQA	SP	SCP	Spot test	Composite result	*K* model	Refinery experience
SN	AH	AEL	0	50/50	U	S	U	U	M	S	S	M	S	S
B1	AH	AL	0	30/70	M	S	S	U	S	S	S	S	S	S
B2	Ratawi	Murban	0	51/49	M	S	S	U	S	S	S	S	S	–
B3	Ratawi	SB	0	53/47	U	S	S	U	M	S	S	S	S	–
B4	Ratawi	AEL	0	53/47	U	S	U	U	S	S	S	S	S	–
B5	Ratawi	AL	0	31/69	M	S	S	U	S	S	S	S	S	–
B6	Ratawi	ML	0	63/37	M	S	S	U	S	S	S	S	S	–
B7	Ratawi	MH	0	62/38	M	S	S	U	S	S	S	S	S	–
B8	RG	Murban	0	24/76	M	S	S	U	S	S	U	S	S	–
B9	RG	SB	0	47/53	U	S	S	U	S	S	S	S	S	–
B10	RG	AEL	0	26/74	M	S	S	U	S	S	M	S	S	–
B11	RG	AL	0	15/85	M	S	S	U	S	S	S	S	S	–
B12	RG	ML	0	38/62	S	S	S	U	S	S	S	S	S	–
B13	RG	MH	0	36/64	S	S	S	U	S	S	S	S	S	–

in a refinery, and the observation of their compatibility/stability during processing is provided in **TABLE 1,** where applicable.

Among all predictions of six different SARA-based methods, the SCP and SP methods were observed to be closer with the *K* model prediction. The maximum deviations of the *K* model results were observed vs. the CII method. Validations of different methods (SCP, spot tests and OCM) are shown in **FIGS. 3–5.** It was observed that all 14 compatibility blends predicted by the prediction model software[a] were in line with the SCP charts (**FIG. 3**). The accuracy of the *K* model was ±1 wt%.

The prediction software[a] determinations of the regions of compatibility/incompatibility for the Ratawi/Saharan blend and the Ras Gharib/Saharan blend were compared with the OCM and spot test methods. The results were a very close match, as depicted in **FIG. 4.** Furthermore, the *K* model is also able to predict the shades of light to heavy crude oils, as the color-intensity shades correspond to variations of the *K* value from low to high.

According to the *K* value, the intensity of the spot test color shade changes, along with the type of crude oil (light and heavy). Higher *K* values are shown as dark colors, with lower *K* values shown as light colors (**FIG. 5A**).

To further validate the predictability, blends were prepared with a *K* value of 0, and the spot tests were observed to be similar in shades, as depicted in **FIG. 5B.**

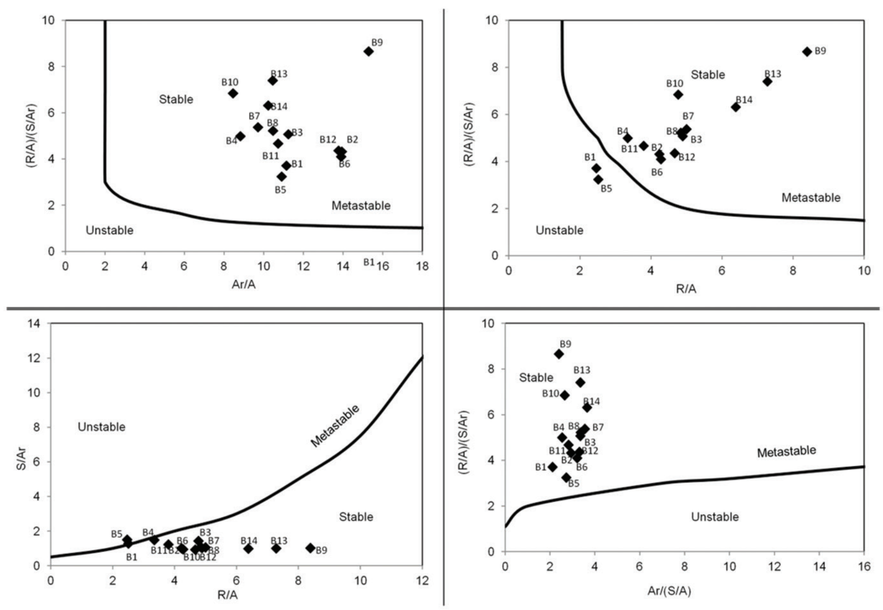

FIG. 3. Validation of the *K* model with SCP method.

This is the level of accuracy displayed by the *K* model to control asphaltene flocculation to precipitation behavior.

Prediction software[a] advantages.

There is no standard practice to check the compatibility parameter in advance; rather, excess antifoulant/chemical dosing is done to prevent incompatibility issues. This is because laboratory-based methods are time consuming. The advantages of the proprietary predicting software are provided in **TABLE 3.**

Impact on refinery operations (equipment, energy and environment).

There is a strong relationship between the *K* model and with the intensity of spot color, desalting performance and fouling behavior, which was further verified through experiments. If the *K* value is positive, then the spot color is darker, and, additionally, desalting is better and fouling is at a minimum. However, if the *K* value is negative, there is a lighter spot color (with an indication of asphaltene flocculation or precipitation), along with poor desalting and high fouling.

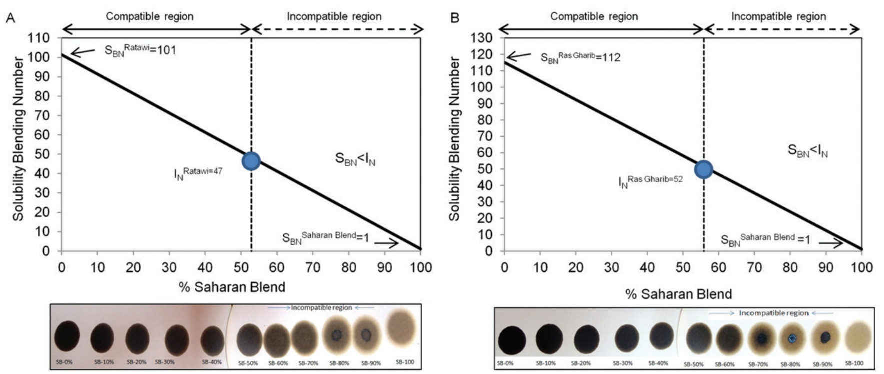

FIG. 4. Comparison of *K* model, OCM and spot test methods of crude oil blend compatibility.

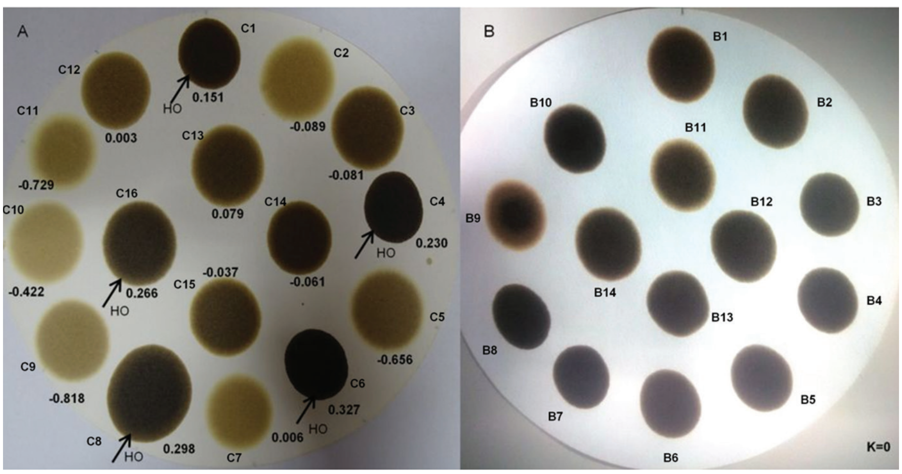

FIG. 5. The *K* model predicting tests of neat crude oils and blends.

TABLE 3. Conventional laboratory testing methods vs. the proprietary prediction software[a]

Conventional method	Proprietary prediction software[a]
A minimum of 3–5 tests are required to ascertain the blend compatibility.	The software predicts the composite results of all nine compatibility test methods, with an accuracy of ±1 wt%
Only two crude oil blends can be tested at one time. To check three or more crude oils, compatibility of the first two must be completed, followed by checking that blend against the third. This method is exhaustive and time consuming.	The model is applicable to any number of known/unknown crude oils for blending compatibility. This is due to the model only needing the four physical parameters of the crude oils being blended.
The results show only the region of compatibility/incompatibility.	The model provides the complete optimized blending solution of multiple crude oils within a compatible region.
Duration of testing	
Two crude oils blending: 1 wk	Two crude oils blending: 2 min–5 min
Multiple crude oils blending: More than 2 wk	Multiple crude oils blending: 1 hr–2 hr

Desalting is a water-washing operation performed to remove salts prior to the crude distillation column. Salt and water content specifications are more stringent because of their negative effect on downstream processes (e.g., corrosion and catalyst deactivation). Incompatible crude oil blends are expected to be problematic for water separation due to asphaltene precipitation, which causes stable oil-water emulsion formation.[10,26]

In the present study, the *K* model-predicted Incompatible Blend A (Ras Gharib 35 wt% and Saharan Blend 65 wt%, with a *K* value < 0) and Compatible Blend B (Ras Gharib 65 wt% and Saharan Blend 35 wt%, with a *K* value > 0) were subjected to desalting experiments. Dehydration efficiency was calculated as the ratio of separated water to the added water after 5 min and 10 min. The dehydration efficiency (water separation) of Compatible Blend B is estimated as 37.78 wt% and 41.12 wt% for 5 min and 10 min, respectively; however, there was no water separation observed with incompatible blend A.

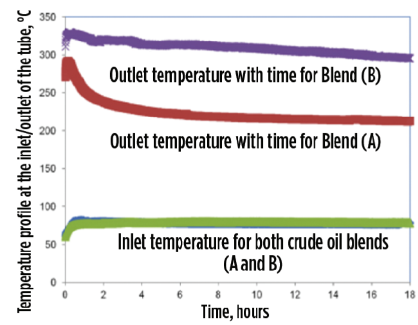

FIG. 6. Fouling profile of compatible vs. incompatible crude oil blends.

The precipitation or deposition of asphaltenes from incompatible crude oil blends is expected during processing.[10,18] In this study, *K* model-predicted Incompatible Blend A (Ras Gharib 35 wt% and Saharan Blend 65 wt%, with a *K* value < 0) and Compatible Blend B (Ras Gharib 65 wt% and Saharan Blend 35 wt%, with a *K* value > 0) were subjected to fouling thermal experiments. In these experiments, the temperature drop profile was measured against duration (**FIG. 6**). The temperature drop of the compatible blends is approximately 34°C; however, the incompatible blends had a high temperature drop of about 59°C. This showed that deposition in Incompatible Blend A is higher vs. Compatible Blend B over the surface of the heating rod; the heat transfer rate reduced significantly, as well. This experimental result is the evidence of a strong relationship between the *K* model and the fouling profile. Therefore, the optimization of blends using the *K* model can be further used for optimizing anti-fouling chemical dosing for fouling mitigation.

The *K* model is expected to improve refinery operations for the desalting and fouling of heat exchangers, which, in turn, helps keep refining equipment in good working order. Asphaltene precipitation typically puts an extra load on the preheat

FIG. 7. Implications of crude incompatibility.

trains (PHTs), which causes a 1°C–3°C decrease in temperature. For context, in a nearly 96,000-bpd refinery, a 1°C improvement in the heat exchanger network (HEN) of the crude distillation unit equals approximately \$1 MM/yr. In addition, to compensate for this, extra fuel firing is required, which typically generates 90,000 tpy of CO_2/°C in the HEN. The implication of asphaltene precipitation on the desalting performance—which is evident from incompatible blends B3, B7 and B8, where there is no water separation—leads to heat exchanger fouling in the downstream processing system and impacts environmental health, as depicted in **FIG. 7.** The adherence to the proprietary prediction software[a] can lead to a significant increase in fuel savings and to a reduction in CO_2 emissions.

Implementation.

Crude oil cost constitutes 85%–90% of a refinery's input costs. For example, if a 350,000-bpd refinery processes crude oil that costs \$1 less, that equates to \$40 MM–\$45 MM in additional profit.[27] While increasing heavy oil processing, incompatibility issues are inevitable. Therefore, the quick and reliable prediction of crude blend compatibility is critical to maintain the good health of refinery equipment and operations.

More than 2,000 crude combinations—involving approximately 200 different crude oils—were analyzed for processing heavy/opportunity crudes in Bharat Petroleum Corp. Ltd.'s (BPCL's) refineries. The prediction software[a] was

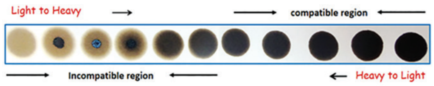

FIG. 8. The prediction software[a] accurately predicts asphaltene precipitation behavior.

used to make quick decisions on feeding a compatible crude mix to maintain the health of refining equipment, while meeting fuels demand. **Note:** It only took 4 wk–6 wk to analyze 2,000 crude mixes, while laboratory studies would have required several months.

Case study: BPCL's Mumbai refinery.

While co-processing a new crude oil at BPCL's Mumbai refinery, the technologists deviated from the prediction software[a], which caused an upset in desalter operations (amperes increased for a longer period due to asphaltene precipitation, and strong emulsion formed with water). To correct this action, technologists used the prediction software to optimize the crude oil blend composition. After 2 hr–4 hr, the desalter was brought back to normal operation. Thereafter, the Mumbai Refinery strongly recommended that the prediction software should be included in the pre-processing plan, so that crude oil blends could be finalized. Since this episode, the prediction software is used regularly in all BPCL refineries.

The prediction software also helped crude-sourcing teams identify which crude oils are best suited for processing and assisted in regular monitoring of as-

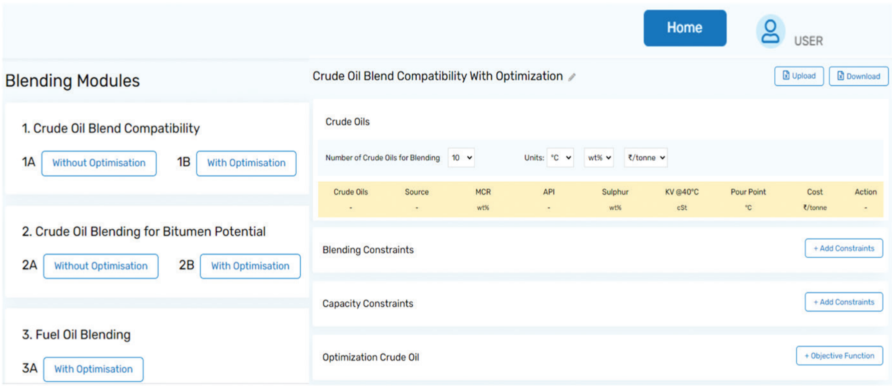

FIG. 9. The prediction software's web page displays software features.

TABLE 4. Prediction software[a] solutions for the selection of heavy oil

SN	Heavy oil	Regular crudes	Crude compatibility	VR compatibility	Recommendations
1	Mexico crude (A)	Arab Medium	Compatible	A < 14 wt%	A < 14 wt%
2		Arab Light	A ≥ 24 wt%	A < 24 wt%	Incompatible
3		Arab Heavy	Compatible	Incompatible	Incompatible
4		AM (60:40)	Compatible	A < 12 wt%	A < 12 wt%
5		AM (70:30)	A > 5 wt%	A < 16 wt%	5 wt% < A < 16 wt%
6		AM (80:20)	A > 9 wt%	A < 22 wt%	9 wt% < A < 22 wt%
7		Arab Ex Light	A ≥ 40 wt%	A < 40 wt%	Incompatible
8		Kuwait	Compatible	Incompatible	Incompatible
9		Upper Zakum	A > 20 wt%	A < 20 wt%	Incompatible
10		Murban	A > 38 wt%	A < 41 wt%	38 wt% < A < 41 wt%
11		Das Blend	A > 35 wt%	A < 35 wt%	Incompatible
12		Basrah Light	A > 8 wt%	A < 9 wt%	8 wt% < A < 9 wt%
13		Iran Light	A > 4 wt%	A < 30 wt%	4 wt% < A < 30 wt%
14		Iran Heavy	Compatible	A < 1 wt%	A < 1 wt%
15		Mars Blend	Compatible	A < 24 wt%	A < 24 wt%

phaltene flocculation to precipitation behavior in refinery operations. The prediction of asphaltene flocculation to precipitation behavior is depicted in **FIG. 8.**

Additionally, the prediction software[a] model predicts the compatibility of intermediate streams within various refining units. This unique feature can predict incompatibility hotspots in various unit operations employed in refineries. The prediction software's solutions have been used by several refinery units to address desalter upsets, the selection of heavy crude combinations, and the compatibility ranking of crude mixes. The predictions and rankings of crude compatibility have been observed in line with field operations.

To enable increased margins, the prediction software guides operators toward enlarging their crude oil baskets, including new and opportunity crude oils for processing. In this case, for example, heavy oil was to be selected for co-processing, which the prediction software predicted that specific crudes were compatible; however, the intermediate streams—especially the vacuum residue (VR)—derived out of the crude mixing process in certain compositions were incompatible, since they could lead to severe fouling in the VR heater/furnace when handling or processing VR in the delayed coker. Based on the VR compatibility, some of the crude oils were avoided for co-processing. In certain instances, the prediction software[a] not only guides the user to a compatible composition, but also advises not to include certain compositions in the crude basket, based on the incompatibility of intermediate streams and refinery configurations. One of the selections of heavy oils—sourced from Mexico—for co-processing with 15 regular crude oils is reported in **TABLE 4.** In this example, the final co-processing decision was primarily influenced by VR compatibility.

Due to its ability for rapid solutions, the *K* model has been used for scheduling healthier feeds for the crude mix processing, including monthly planning and real-time monitoring of asphaltene precipitation behavior. The continuous usage of the prediction software resulted in an increase in refining margins of $0.15/bbl–$2/bbl vs. regular crudes.

Prediction software[a] modules and features. The prediction software package offers different modules that provide a variety of options to meet refinery blending needs. This software package has three modules, as depicted in **FIG. 9.** The first module is for crude oil blend compatibility and has two submodules. Module 1A is used to predict crude blend compatibility for known blend compositions. Module 1B is used to predict the most optimum and compatible crude oil blends that can be achieved using a set of crude oils. The prediction software provides an option for blend optimization of up to 10 crude oils. This option considers different blending constraints, such as blend compatibility, blend viscosity, blend pour, blend acidity, blend nitrogen, blend sulfur and total distillate yield, and crude oil availability.

Similarly, Module 2 is used to indicate possible bitumen within blends. Module 2 also has two submodules. Module 2A can check if a given crude oil blend composition has the potential for bitumen. Module 2B is used for estimating the optimum blend composition of crude oils with bitumen potential.

The third module is used for fuel oil blending, which enables the optimization of cutter stock and also minimizes the cost of production for a fuel oil of a certain specification.

Takeaway. The authors' company's proprietary crude compatibility software[a] provides accurate prediction of crude compatibility within minutes vs. weeks. This software has been validated with the composite results of nine different compatibility tests available in literature for accurate compatibility prediction.

The ability to predict crude oil compatibility can significantly improve refining margins, especially for refiners that work with large crude mix parcels. For cases where 4–10 crudes are likely to be blended, predicting compatibility will be a necessity and not a luxury. The prediction software also guides refiners for optimal fuel oil blending and crude blending for bitumen production. Most importantly, this will assist refiners to process crude mix parcels with a greater number of constituents to increase profitability.

Refiners can use the software to improve equipment life, ensure smooth operations, increase energy savings and lessen environmental impact. This software has the capability to significantly contribute to the world if refineries adhere to minimizing asphaltene precipitation. HP

NOTE

[a] BPCL's K Model

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HP Special Focus

Digital Technologies

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Why robotic process automation is quickly becoming a must for downstream companies

For the refining operation that wants to relieve its planning team of mind-numbing manual data-management tasks, there are bots for that. For the fuel distributor that wants to complete scheduling tasks with greater speed and precision and with minimal human involvement, there are bots for that and many other tasks, too.

Bots are quickly becoming fixtures across the energy landscape as more producers, refiners, distributors and other participants in the hydrocarbon supply chain turn to a technology known as robotic process automation (RPA) as an important step toward becoming an intelligent, data-informed enterprise with the side benefit of unlocking additional value through increased efficiency, productivity, reliability and predictability. Already widely used in industries like banking, finance, insurance and even utilities to help companies boost efficiency through automation and manage vast volumes of data from disparate sources, RPA is gaining a stronger foothold in the oil and gas business, which has a long history of process automation. According to the author's company, 40% of oil and gas companies pursuing efficiency improvements utilize increasingly digital technologies like RPA as part of their transformation initiatives. To power that transformation, downstream energy companies are using RPA in tandem with technologies like the Internet of Things, predictive analytics, machine-learning and artificial intelligence—the building blocks of an intelligent enterprise.

As one of those building blocks, RPA can help energy companies capture new efficiencies and improve productivity across their operations and supply chains simply by automating tedious, time-consuming and often error-prone manual processes. These automation capabilities are critical to improving resilience and maintaining margins during volatile oil and gas market cycles.

Integrating RPA into an energy business does not mean replacing human beings with actual robots. It is about using intelligent software to perform repetitive tasks, freeing people for pursuits that add value to the business. Guided by a set of user-defined rules, RPA uses programmable, machine-learning-driven bots to automate tasks. Some bots need prompting and interaction with a human user ("attended" bots) to perform a programmed task. "Unattended" bots are trained to work autonomously without human input. Today's generation of bots actually "watch" users perform processes, intuit the logic behind them and determine how to do the same tasks themselves. RPA interacts with individual computer systems the same way a human user would, with no complex integration or coding required. Moreover, energy companies can access RPA as a cloud-based service without an expensive capital outlay.

The ease with which RPA can automate certain key business functions makes it invaluable to a downstream energy company, particularly in the areas of planning and scheduling, where a large percentage of the data inputs (typically 60%–80%) tend to require time-consuming and error-prone manual intervention and processing in a variety of formats: Excel, PDF, email, etc. As more downstream energy companies are discovering, RPA can save significant amounts of labor and reduce costly accounting, planning and scheduling errors, all of which benefit the bottom line. The myriad tasks that RPA bots can perform for an energy company include:

- Trading confirmation processing
- Trading settlement and invoicing
- Upload of general ledger entries
- Upload of supplier invoices
- Payables scanning, matching and processing
- Order confirmation
- Production order operation confirmation
- Supplier invoice status checks.

It is no wonder bots are becoming so popular. Two examples of how energy companies are putting them to work in a downstream setting are discussed here.

To streamline lead-to-cash. In a volatile price environment companies must have the ability to quickly lock in prices and promptly collect payment. RPA can streamline the lead-to-cash process, using bots to correct missing master data to transact deals without disruption. Meanwhile, financial-related bots can automate certain time-consuming processes, such as invoice status

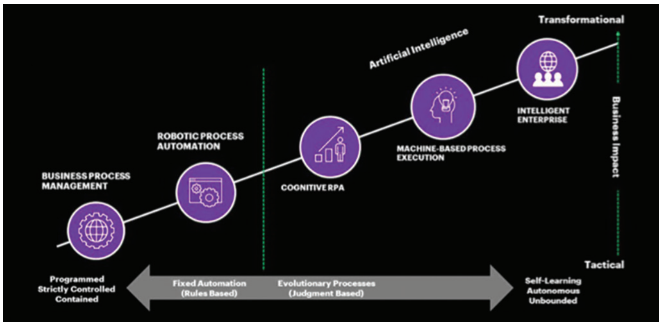


FIG. 1. Developing capability along intelligent automation spectrum

checks, with the option to alert non-paying customers that their payments are due. Additionally, bots can automatically upload invoices to the cloud, enable smart accruals from spreadsheet to accounting system, and create sales inquiries via screen scraping.

To improve vessel and/or berthing planning. Having real-time visibility into transportation vessels' current and planned locations is a must. Instead of creating vessel or berthing plans based on details from a haphazard collection of emails and information from multiple sources—tanker plan, notice of readiness, cargo loading completion date and time, bill of lading, expected time of berthing at port, expected time of sailing off at a discharge port and all the other moving parts—bots can streamline the process through automation, while also ensuring real-time updates are made to tanker plan reports. The result is a sharp reduction in manual effort, a seamless vessel planning process and significantly less pre-berthing time. **FIG. 2** shows how this process might unfold with multiple bots on the job.

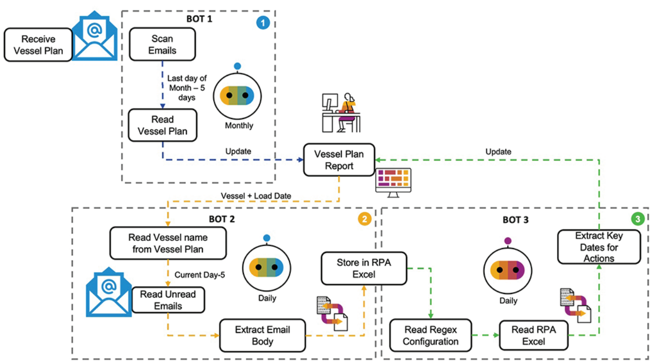
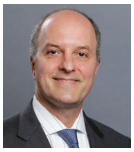


FIG. 2. RPA automation for vessel planning. Source: Prasanth Padmanabhan Menon of SAP.

Ultimately, all the moves these bots make behind the scenes, i.e., planning, scheduling or some other aspect of an operation, contribute to building a more efficient and intelligent downstream enterprise.

Taking advantage of RPA is merely a matter of identifying the manual activities across the business that could be automated—there likely are many—then finding an RPA provider with the right mix of pre-built automations that can be readily integrated into the enterprise resource planning system. Once those boxes are checked, it is time to sit back and watch your bots in action. HP



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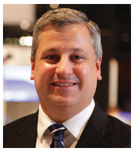

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Debottlenecking of an extractive distillation column for aromatics recovery: Effect of feed pre-vaporization

Extractive distillation is one of the most efficient techniques to separate aromatic and non-aromatic hydrocarbons by increasing the relative volatility of the mixture in the presence of a solvent. Extractive distillation is attractive because of its low capital and operating costs, operational ease and flexibility. It has many applications within the refining and petrochemical industries for the recovery of high-purity products, including benzene, toluene and xylene (BTX), butadiene, isoprene and styrene. In the hydrocarbon processing industry, it is commonly used for treating light reformate in refineries. The performance of the extractive distillation column can be restricted for several reasons, ranging from operational to mechanical factors. This article details a case study of high pressure drop observed across the extractive distillation column, the operational reasons for it and a cost-effective solution to this challenge. A significant increase in column pressure drop can lead to flooding. The loss of stages due to flooding adversely affects separation and product quality.[1]

FIG. 1 shows a simplified schematic flowsheet for an extractive distillation unit separating aromatics from the hydrocarbon stream. The non-aromatics exit as raffinate from the top of the extractive distillation column. The solvent and aromatics phases are routed to the solvent recovery column, where solvent is separated from aromatics by conventional distillation. The lean solvent reenters the extractive distillation column by exchanging heat with the feed in the preheater.

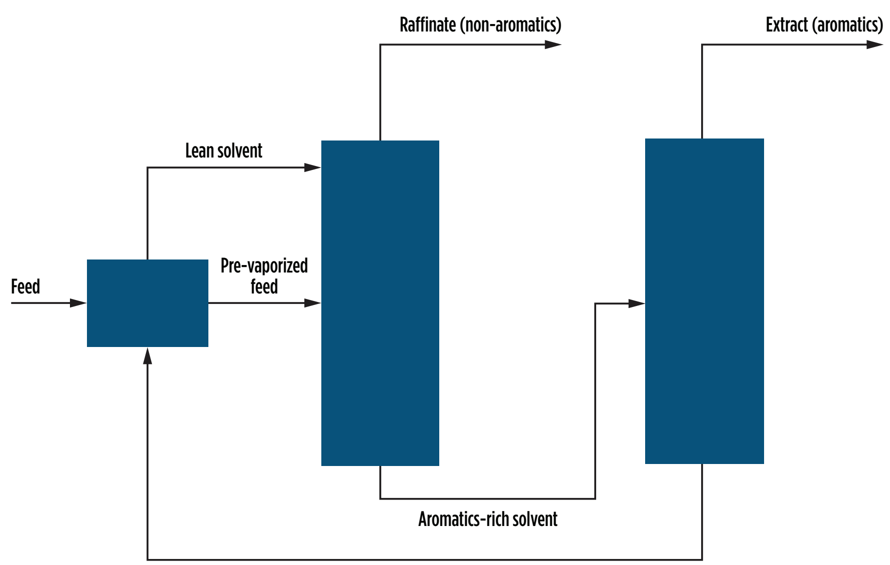

FIG. 1. Simplified schematic flowsheet of the extractive distillation unit for the aromatics separation process.

The operation of the extractive distillation column was constrained due to high pressure drop, especially in the bottom section below the feed tray. To gain insights into the problem, the unit was first simulated using a commercial process simulation software, and the preheat train was simulated using a commercial heat exchanger simulation software.

The issue of high pressure drop was addressed by increasing feed pre-vaporization. Many studies have discussed the impact of feed vaporization on column reboiler duty. However, this study focuses on the impact of extractive distillation feed vaporization on various column parameters, including pressure drop, reboiler duty and aromatics separation efficacy. The process simulation results for this case study showed that increasing the feed pre-vaporization can help to reduce pressure drop across the extractive distillation column, thus saving the reboiler duty without significantly affecting aromatics recovery.

This simple philosophy of increasing feed pre-vaporization to tackle the challenge of high pressure drop in columns can be tested in similar plant problems in other operating plants, considering all the other constraints present in this respective operating plant, some of which are conveyed in this article.

Ideology. The adequacy of the aromatics extractive distillation column is limited by high pressure drop. If altering process parameters does not help, then this

issue can be addressed by changing the column internals, which may not always be an economically viable solution. One of the solutions could be to increase the feed pre-vaporization. This would help to reduce the reboiler duty, along with the vapor-liquid traffic in the bottom section of the extractive distillation column, thereby decreasing pressure drop across the column. The degree of pre-vaporization is dependent on the percentage of non-aromatics in the feed. This possibility was checked by analysis of the extractive distillation unit preheat circuit, using a simulation. The extent of feed pre-vaporization is primarily governed by the:

- Maximum feed temperature attainable across the feed preheater in the extractive distillation heat exchanger circuit (the higher the feed temperature, the greater the extent of pre-vaporization)
- Ability of the feed line to sustain the flow regime with increased vapor flow
- Acceptable slippage of aromatics and non-aromatics in the product streams, as per the operating plant's specifications.

Vaporizing the feed ensures that light boiling components go up the feed tray in the vapor phase. This avoids the light non-aromatic components entering the bottom section of the column, which saves reboiler duty. The reboiler duty goes down to maintain heat balance inside the extractive distillation column, and eventually decreases the vapor flow through the trays between the feed tray and the bottom tray. This helps to reduce the pressure drop in the stripping section of the column. Therefore, feed pre-vaporization might be a practical solution in case the column is limited by flooding.

However, increasing the feed vaporization does not always improve the overall separation efficiency of a distillation unit. Excessive feed temperatures can cause a significant amount of flash of heavy key components at the distillation column feed zone. The extent of pre-vaporization will depend on the allowable specifications of aromatics slippage in the raffinate stream of the extractive distillation, as well as non-aromatics slippage in the extract stream of the solvent recovery column. In addition, the capability of the feed line to endure the increased vapor fraction must be evaluated. Personnel must look at the effect of feed pre-vaporization on the operation of downstream units like the solvent recovery unit (SRU).

> **Increasing feed pre-vaporization results in lowering vapor-liquid traffic in the stripping section of the extractive distillation column, which leads to reduced pressure drop and reboiler duty, thereby debottlenecking the column.**

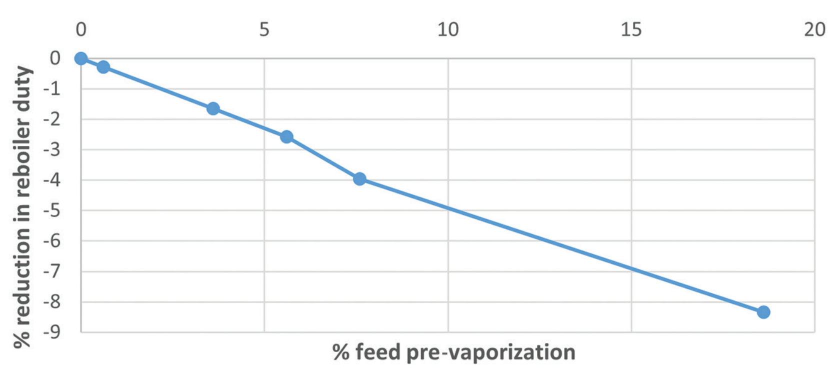

FIG. 2. The impact of feed pre-vaporization on the reboiler duty of the extractive distillation column.

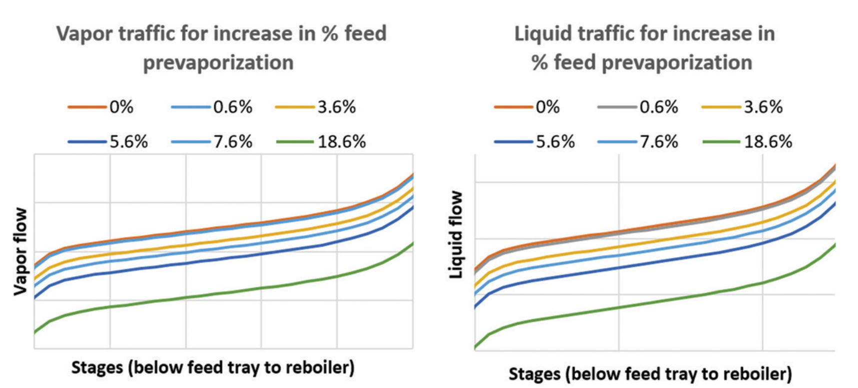

FIG. 3. Trends of V-L load below the feed tray with an increase in feed pre-vaporization percent.

Approach. The following methodology was used for the case study:

- Perform a simulation of extractive distillation and solvent recovery columns using commercial simulation software and an appropriate thermodynamic property model
- Check the possibility of feed pre-vaporization, with the help of simulation
- Perform the extractive distillation column hydraulic study (pressure drop studies) by increasing feed pre-vaporization, and study its impacts on reboiler duty and on aromatics and non-aromatics separation.

Analysis of the extractive distillation unit preheat circuit. To check the possibility of feed pre-vaporization and the maximum reasonable feed temperature across the feed preheater, it was important to analyze the extractive distillation column's preheat circuit to check the adequacy of all heat exchangers in the circuit. Depending on the preheater duty margin and flashing conditions, the authors used simulation models to determine the feed pre-vaporization percent by increasing

feed temperature. The authors compared the results for cases with 0%, 0.6%, 3.6%, 5.6%, 7.6% and 18.6% of feed pre-vaporization. Reboiler duty and specifications of the end products were checked for each scenario. In addition, the transition in flow regimes with vaporization was studied to inspect the feed line's effectiveness.

Impact of feed pre-vaporization on reboiler duty. The impact of feed vaporization to offload reboiler duty in distillation columns is discussed in literature.[2] Likewise, the authors' simulation for the case study indicated a reduction in reboiler duty with increasing feed pre-vaporization (**FIG. 2**). A further reduction in reboiler duty from the existing level would depend on the prospects of pre-vaporization due to the reboiler margin and the preheating of feed to the extractive distillation unit.

Impact of feed pre-vaporization on pressure drop. The increase in feed pre-vaporization decreases vapor-liquid (V-L) traffic in the extractive distillation column, especially in the stripping section (below the feed tray). Using simulation software, stagewise V-L profiles were generated. The decreasing trends for vapor and liquid load below the feed tray to the reboiler for each percent vaporization are shown in **FIG. 3.** The V-L load is consistent above the feed tray. The impact of pre-vaporization on pressure drops in the stripping and rectification sections of the column, and on pressure drops of the entire column, is depicted in **FIG. 4.** It is evident that, with increased feed vaporization, pressure drops in the column could be reduced further.

Impact of feed pre-vaporization on aromatics separation. The increase in feed pre-vaporization might affect separation efficiency of the extractive distillation column. When it comes to extractive distillation separation, for the raffinate phase, the loss of aromatic hydrocarbon components must be monitored. Conversely, in the extract phase, non-aromatics content is an important criterion, since it affects final product quality. The impact of feed preheating on both parameters is depicted in **FIG. 5.**

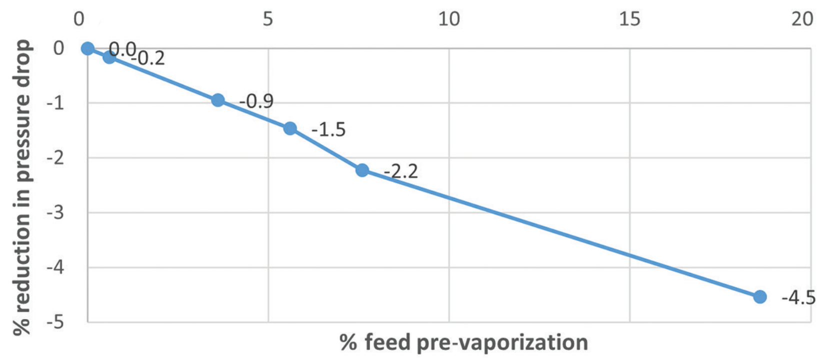

FIG. 4. The impact of feed pre-vaporization on pressure drop of the extractive distillation column.

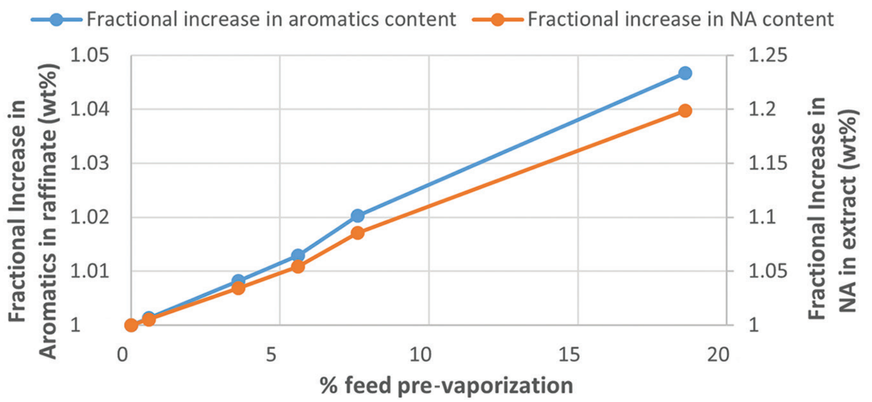

FIG. 5. The impact of feed pre-vaporization on aromatics in raffinate and non-aromatics in extract.

TABLE 1. The effect of feed pre-vaporization on product specifications

Cases	Base/Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Feed pre-vaporization, %	0	0.6	3.6	5.6	7.6	18.6
Percent increase in feed throughput	0	0.2	0.9	1.5	2.2	4.5
Column parameters	**Base/Case 1**	**Case 2**	**Case 3**	**Case 4**	**Case 5**	**Case 6**
Feed flow, kg/hr	X	(1.002*x)	(1.009*x)	(1.015*x)	(1.022*x)	(1.045*x)
Pressure drop, kg/cm^2	y	y	y	y	y	y
Aromatics content in raffinate, wt%	a	(1.001*a)	(1.008*a)	(1.013*a)	(1.02*a)	(1.05*a)
Non-aromatics in extract, wt%	b	(1.01*b)	(1.03*b)	(1.05*b)	(1.09*b)	(1.2*b)

Aromatics slippage in raffinate increases due to increased pre-vaporization of the feed. Similarly, the slippage of non-aromatics also increases. However, it should be confirmed that the loss is well within desired product specifications.

Scope to increase feed throughput. In the studies and feed conditions, the authors found that increasing feed pre-vaporization can help to reduce the pressure drop across the extractive distillation column. This provided the authors with a scope to increase feed throughput by the same fraction as pressure drop decreases, thus maintaining base pressure drop without significantly affecting separation efficiency. For example, at 7.6% feed vaporization, the pressure drop decreases by 2.2% (FIG. 4). Therefore, the feed throughput can be increased by 2.2%, thus debottlenecking the column. The same principle can be applied to other pre-vaporization cases by retaining base pressure drop. The process constraints and product specifications must be monitored for any such alteration.

TABLE 1 considers a base case depicting some column parameters using alphabetical variables. The effect of increasing feed throughput for each pre-vaporization case on column parameters is also represented. It should be noted that the operation of the SRU was undisturbed, and it did not hit any operation or quality constraints due to increased pre-vaporization of the feed.

Takeaway. For this case study, with an increase in feed pre-vaporization, the pressure drop across the extractive distillation column and reboiler duty was found to decrease. The degree of pre-vaporization will depend on the competence of the extractive distillation heat exchanger circuit and desired product specifications.

This case study provides a cost-effective and accessible solution that can be further extended for similar plant challenges where extractive distillation columns are bounded by high pressure drops. It should be noted that the operation of the downstream unit (i.e., the solvent recovery column) was undisturbed by this modification. HP

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Impact of biofeed retrofits, co-processing on refinery amine units, SWSs and SRUs—Part 2

Producing diesel with a portion of biologically sourced carbon is being done at an increasing number of conventional crude oil refineries. This is commonly achieved by co-processing biofeedstocks in refinery hydrotreaters and FCCUs or through the installation of a dedicated biofeedstock hydrotreater to produce commercial biodiesel products. Generally, refineries are looking to technologies that allow them to easily incorporate biofeedstocks into their existing infrastructure. Part 1 of this article, published in the January issue of *Hydrocarbon Processing*, described the impacts that this incorporation has on a refinery's existing amine and SWS units. Part 2 will detail the impacts on the SRU, operational and design options available to manage those impacts, and a specific case study.

Impacts on the SRU. The amine acid gas (AAG) from an amine unit and sour water acid gas from an SWS unit are fed into one or more SRUs. FIG. 8 shows the typical process flow diagram for a two-stage SRU. In the reaction furnace, a portion of the H_2S is partially oxidized to SO_2, which then reacts with the remaining H_2S to form elemental sulfur over the thermal and catalytic stages of the process. A detailed description of the process and chemistry is provided in literature.[18] The co-processing of biofeedstock will change the amount and gas composition to the SRU. This will be mostly an increase of CO_2 and a reduction of H_2S. Processing the changed feed is unlikely to be a problem since most SRUs can handle changes in composition and load to the unit. While this means that the existing equipment may be adequate, it does require careful review before committing to biofeedstock processing.

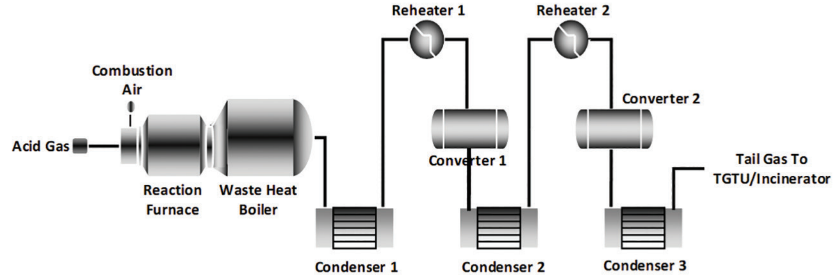

FIG. 8. Process flow diagram of a two-stage modified Claus unit.

The impact of the extra CO_2 on the SRU is limited to the thermal stage (reaction furnace). The thermal stage not only oxidizes H_2S and thermally converts it into elemental sulfur, but it also helps to destroy feed contaminants such as hydrocarbons, benzene, toluene, ethylbenzene and xylene (BTEX) and ammonia. If not destroyed, these contaminants can cause operational challenges such as catalyst damage and plugging in the downstream equipment due to salt formation (FIG. 9). The additional CO_2 will cool the bulk gas temperature in the reaction furnace, which could reduce contaminant destruction. With only hydrocarbons and BTEX in the feed, the minimum target temperature is 1,050°C. When sour water acid gas (SWAG) is one of the feeds, the key requirement is to ensure good destruction of ammonia, meaning that the process gas temperature in the furnace is recommended to be at least 1,250°C.[19]

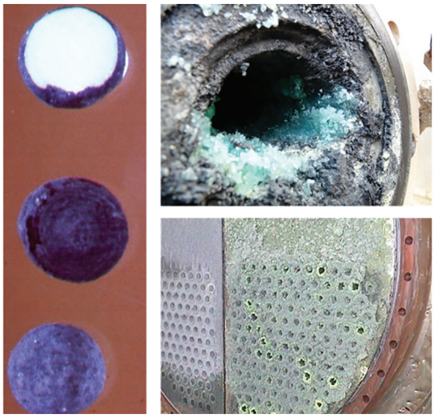

FIG. 9. Catalyst deactivation by contaminants (left) and by plugging in equipment (right).

The ability to achieve the target temperature will be reviewed for the two most common thermal stage configurations (FIGS. 10A and 10B): the straight-through process (single burner, single chamber) and the split-flow process (single burner, two chambers). The advantages and disadvantages of both configurations have been well documented in literature.[19]

Straight-through furnace. In the

straight-through (one burner, one chamber) configuration, the reaction furnace temperature is strictly a function of the process chemistry. The extra CO_2 from the biofeedstock only cools the furnace. When processing SWAG, if the resulting furnace temperature is expected to be less than the 1,250°C target value, then more ammonia will leave the furnace and increase the risk of ammonia salt formation and plugging (**FIG. 9**) in the downstream waste heat boiler and subsequent equipment.

Although increasing the furnace temperature is possible by warming the feed gas(es) and combustion air with preheaters, there is a considerable capital investment in addition to the extra operating cost and maintenance complications. Alternatively, the temperature can be increased by co-firing with some type of fuel gas or by oxygen enrichment of the combustion air stream.

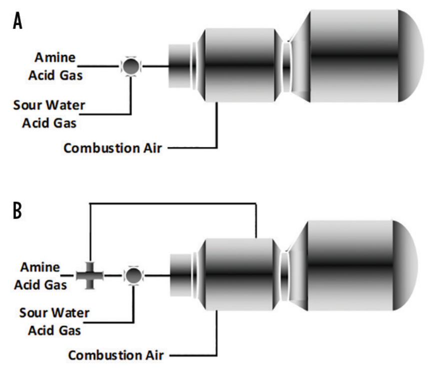

FIG. 10. Views of a straight-through furnace configuration (A) and a split-flow furnace configuration (B).

Fuel gas is also used in the SRU for startup, hot standby and shutdown operations, as a reliable source of energy to manage the furnace temperature. The fuel may be natural gas, refinery fuel gas, hydrogen, etc. Whichever gas is used, it is critical that it is of constant composition to maintain good process control. Large fluctuations in the gas composition, and, thus, combustion air demand, will interfere with control of the $H_2S:SO_2$ ratio target, meaning lost recovery efficiency and other possible dangers. If continuous co-firing is selected as the heating source, the design of the burner must first be reviewed. Next, ensure the SRU's maximum throughput or hydraulic load can handle the extra load from co-firings, as co-firing will add a large amount of inert gas to the SRU. The added dilution also reduces the sulfur recovery efficiency, which may be unacceptable. Finally, the CO_2 emissions of the facility will increase, which may be a concern in some locations.

A cleaner and better way to improve the furnace temperature is to increase the oxygen concentration of the combustion air. Oxygen enrichment has been used successfully in SRUs that are processing lean (i.e., low concentration H_2S and more CO_2) acid gas to achieve the desired furnace temperature. However, this requires some modification to the system. The target oxygen concentration will determine the extent of the changes required—low-level oxygen is quite a simple addition, while greater concentrations (up to 100% oxygen) require a special burner and materials with a dedicated oxygen lance.[20] The source and cost of oxygen must also be considered. It is important to check the turndown capability of the burner, since the addition of oxygen to the process increases not only the flame temperature, but also the speed of combustion. Usually, systems equipped with oxygen enrichment require more throughput to ensure that the burner operates in a safe regime to protect the metal and refractory components.

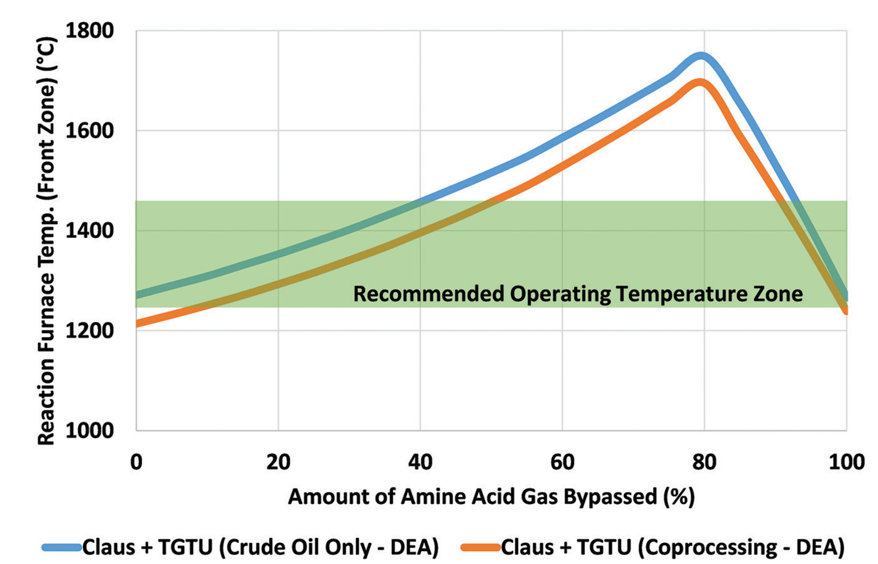

FIG. 11. Acid gas bypass rate vs. front-zone reaction furnace temperature in a SWAG-processing SRU.

Split-flow furnace. In the split-flow process, the SWAG containing the ammonia is processed directly through the main burner with a portion of the AAG. The balance of the AAG is sent to the rear zone of the furnace and is used to control the temperature in the front zone (**FIG. 11**). The CO_2 content of the SWS, especially with biofeedstock processing, may reduce the effectiveness of the bypass design since this increases the gas volume without adding heat of combustion. If the target temperature cannot be achieved with the split-flow design alone, then co-firing or oxygen enrichment can be used.

The rear zone of the furnace must also be hot enough to destroy the BTEX in the bypassed acid gas. This requires a minimum burn temperature of 1,050°C,[21] but this threshold was determined using straight-through reaction furnaces only with sufficient residence time, which is another serious process consideration. Conservatively, hotter operation of the second zone is recommended, but BTEX destruction can only be safely confirmed by field analytical testing.

The impact of biofeed co-processing on SRU reaction furnace temperature, and the effect of mitigation methods (such as natural gas co-firing, oxygen enrichment and operation with a split-flow furnace) are discussed further in this article's case study.

Although more CO_2 in the feed(s) will form more carbonyl sulfide (COS) in the reaction furnace, losses associated with this species can be managed in the first Claus converter with temperature control, catalyst choice and catalyst activity. Similarly, more CS_2 is formed when co-firing, which is also managed in the first converter. For the many facilities that have an amine-based TGTU after the Claus section, virtually all losses from

COS and CS_2 will be eliminated, as these species will hydrogenate to H_2S and be recycled to the front of the SRU.

Impacts on the TGTU. Today's strict SO_2 emissions requirements are beyond the capabilities of the Claus process. The tail gas cleanup unit is a common addition to adhere to the new guidelines. Tail gas units include processes such as sub-dewpoint, direct oxidation, hydrogenation with amine absorption and stack SO_2 recovery. The Claus tail gas hydrogenation process followed by amine absorption—called a TGTU—is the most popular choice when more than 99.5% sulfur recovery efficiency is required.[22] The impact of biofeed co-processing on TGTUs is important not only due to its prevalence in refineries, but also because the acid gas from the regenerated amine is recycled to the reaction furnace. Alternate tail gas processes, such as sub-dewpoint or direct oxidation, are not discussed in this article because the changes brought about from biofeed co-processing are minimal, apart from those that have been discussed in the SRU section.

In a TGTU, the Claus tail gas is first reheated and processed in the hydrogenation reactor in which all oxidized sulfur species (SO_2, COS and CS_2) and the remaining elemental sulfur are hydrogenated or hydrolyzed to H_2S on the surface of cobalt-molybdenum catalyst. The hydrogenated process gas is then cooled through an exchanger and water quench column and then treated in the amine absorber. The rich amine is pumped to the regenerator, where the acid gas species are stripped from the solvent and recycled back to the SRU reaction furnace. The TGTU amine is typically a selective amine (most commonly MDEA) that is optimized to minimize CO_2 absorption, which minimizes the CO_2 recycled to the SRU.

With biofeed co-processing, the composition of the tail gas of a well-operated Claus unit is expected to be like the original case, except for an increase in CO_2. While the extra CO_2 will pass through the hydrogenation section of the TGTU, the increased CO_2 in the feed will increase its pickup by the amine, and the rich amine loading and CO_2 recycle will, therefore, also increase. To maintain optimum performance, the amine circulation rates or amine strengths will need to be adjusted or a switch to a more selective amine chemistry may be required.

Installing acid gas enrichment (AGE). To maintain hydrotreater catalyst activity, a certain amount of sulfur is needed in the feedstock. This is primarily of concern in the HVO production route, where there is not always sufficient sulfur naturally in the hydrotreater feed to do this. This can be overcome by the addition of a sulfiding agent, such as dimethyl disulfide (DMDS), to the hydrotreater feed; however, continuous addition can be expensive, and sulfiding agents are often difficult (and unpleasant) to store and handle. One solution is to install an AGE unit that can concentrate the H_2S from the amine process and recycle it to the hydrotreater, keeping the catalyst active. The installation of an AGE unit design is a decision that can be advantageous for some facilities. More facility details on where AGE systems have been installed, as well as the rationale for their installation, are provided in literature.[23,24]

CASE STUDY: FICTIONAL EXPERTS REFINERY

The fictional 100,000-bpd Experts Refinery has been built for maximum diesel yield, and generally processes high-sulfur crudes. The process conditions are loosely based on the field test results of several operating refineries optimized for diesel yield. The SWS and amine plants were simulated on a proprietary gas treating simulation tool[a], while the SRU and TGTU were simulated on a process simulation software[b]. The authors have found that these two programs yield the best accuracy of the respective simulated units.

The Experts refinery has two diesel hydrotreaters (cracked and straight-run) and co-processes 10% biofeedstock in both. This has an impact on the treating units in several ways. A comparison of normal and co-processing operation is provided in **TABLES 4–7.**

The key parameters for the SWS unit are shown in **TABLE 4.** In this case, a stripped water specification of 50 ppmw ammonium ions is desired. To meet this specification with biofeed co-processing, stripping steam was increased by 23%. Much of this steam was required to process the additional water volume. The additional steam also required more reflux condenser capacity with the associated condenser duty to reduce the SWAG temperature to 85°C, which had increased by 25%. While many facilities will not have this ability, this duty can be offset to an extent by allowing the overhead tempera-

TABLE 4. SWS key parameters, crude only vs. co-processing operation

Parameter	Crude oil only	Co-processing
Sour water volume, m^3	70	85
Sour water CO_2, mg/l	1,500	3,500
SWAG CO_2, vol%	3.2	8.6
SWAG flow, tpd	42	47
Stripping steam, kg/hr	14,200	17,500
Condenser duty, MW	5.6	7

TABLE 5. Main amine system key parameters, crude only vs. co-processing operation

Parameter	Crude oil only, DEA	Co-processing, DEA	Co-processing, MDEA
DHT hydrogen CO_2, %	0	0.5	0.5
DHT hydrogen H_2S, %	1.5	1.4	1.4
Non-DHT H_2S, kmol/hr	107	107	107
Non-DHT CO_2, kmol/hr	18	18	9
H_2S-rich loading, m/m	0.32	0.32	0.25
CO_2-rich loading, m/m	0.04	0.07	0.02
Stripping steam, kg/hr	23,000	24,000	20,000
Condenser duty, MW	5.4	5.7	5.2
AAG H_2S, vol%	84	78	86.9
AAG CO_2, vol%	10.5	16.5	7.87
AAG flow, kg/hr	5,650	6,120	5,470

ture to increase, provided there is hydraulic capacity in the downstream SRU to handle the additional water vapor in the SWAG stream. In practice, capacity in the water feed-effluent exchanger should also be verified prior to co-processing.

With less residence time in the upstream separator vessels on the diesel hydrotreaters (due to significantly increased water being generated in these units), there will be more hydrocarbons in their sour wastewater. A liquid-liquid coalescer may be required on this stream. The impact of additional organic acids in the sour water due to co-processing (e.g., phosphates) was not considered. In practice, if the level of additional organic acids is significant, caustic can be injected to neutralize them.

The refinery's main amine system uses DEA, which is common in many refineries. **TABLE 5** details the impact of co-processing on key parameters. The table includes a case where the DEA solvent is replaced with an MDEA solvent, which preferentially absorbs H_2S over CO_2.

Typically, in amine units, the limits on the exchanger duty design are reached before system hydraulics. The extra CO_2 absorbed by the DEA, with co-processing, requires more reboiler duty (50 psig stripping steam in this study) to regenerate the amine and, hence, additional condenser duty in the top of the stripper column. In this case, 4%–5% more reboiler and condenser duty were needed. This should not be a problem for systems not near capacity; however, many older amine systems are already operating at capacity—often because utilities have not been expanded at the same rate as refinery capacity—and, for these systems, some equipment capital investment may be required. Alternatively, the solvent could be changed to MDEA.

Switching to an MDEA solvent has advantages in terms of energy consumption and improved H_2S:CO_2 ratio in the AAG, which limits the need for downstream modifications of the SRU. However, this is not the only consequence of changing to MDEA, and the effects on other units in the refinery need to be considered before a system-wide conversion.

The refinery's SRU, which processes both SWAG and AAG, was modeled under two different cases: a two-stage Claus only and a two-stage Claus with an amine-based TGTU. The results are provided in **TABLE 6.** All cases used DEA in the upstream main amine system. The most obvious change in the SRU is the increased CO_2 concentration in the feed acid gas when co-processing. The volumetric flow through the SRU increased by 2%–3%, which is minor. The Claus section's sulfur recovery efficiency declined by 0.1%, mainly due to the decrease of H_2S in the acid gas and the cooler reaction furnace.

As expected, an increase of CO_2 in the feed reduced the reaction furnace temperature. It should be noted that the temperature presented in **TABLE 6** is an adiabatic temperature and does not consider heat losses, which means that the actual temperature of the furnace during co-processing is expected to be less than 1,250°C. As discussed in the SRU section, maintaining the reaction furnace front zone temperature at, or hotter than, 1,250°C is critical for complete destruction of feed contaminants.

If the reaction furnace is equipped with a two-zone chamber, then the acid gas bypass fraction to the rear zone can be increased to maintain the front-zone hotter than 1,250°C. In this case study, the bypass fraction must be between 10%–50% of the inlet acid gas flow for maximum temperature to destroy the contaminants (**FIG. 11**). The upper limit is based on typical refractory thermal limitations.

As discussed in the SRU section, if the furnace has a straight-through design, the furnace temperature can be

TABLE 6. SRU key parameters, crude only vs. co-processing operation

	Modified Claus unit		Modified Claus unit + TGTU	
Parameter	Crude oil only	Co-processing	Crude oil only	Co-processing
Acid gas flowrate, kg/hr	5,650	6,120	5,650	6,120
Recycle acid gas flow, kg/hr	–	–	553	873
Total acid gas flowrate, kg/hr	5,650	6,120	6,203	6,993
SRU total feed H_2S, vol%	68.5	64	66.5	60.7
SRU total feed CO_2, vol%	8.3	14	11.3	18.6
SRU total feed NH_3, vol%	12.8	11.6	12.1	10.7
Reaction furnace temperature, °C	1,299	1,253	1,271	1,214
CO_2 in Claus tail gas, °C	2.6	4.6	3.6	6.5
Claus tail gas flowrate, Nm^3/hr	14,270	14,484	14,663	15,066
Claus sulfur recovery, %	97.39	97.3	97.25	97.13

TABLE 7. TGTU amine section key parameters, crude only vs. co-processing operation

Parameter	Crude oil only	Co-processing (no change in TGTU)	Co-processing (TGTU optimized—high circulation)
Absorber inlet flowrate, kgmol/hr	455	475	475
Absorber inlet CO_2, %	6.46	10.98	10.98
Absorber inlet H_2S, %	1	1	1
Absorber outlet H_2S, ppmv	139	500	150
Absorber CO_2 slip, %	70	75	74
Amine flowrate, m^3/hr	41	41	45
Amine strength, wt%	40	40	40
H_2S rich loading, m/m	0.036	0.036	0.033
CO_2 rich loading, m/m	0.061	0.094	0.089
Stripping steam, kg/hr	2,620	2,620	3,100
Condenser duty, MW	0.43	0.38	0.54
AAG H_2S, vol%	32.45	24.14	24
AAG CO_2, vol%	62.21	70.59	70.72

managed with oxygen enrichment or co-firing. FIG. 12 shows the reaction furnace temperature profile for a combined acid gas feed (main amine system and TGTU recycle streams), which contains 22.4% CO_2 because of co-processing. In this case, the model shows that the minimum required temperature of 1,250°C in the reaction furnace can be easily achieved with low-level oxygen enrichment without other changes.

In the case of co-firing, it was assumed that the fuel source is a pure methane natural gas. Simulation shows that the introduction of the co-processed acid gas increased the volumetric flow in the SRU by 2.3%, which should be accommodated in the existing plant. However, every 5 kmol/hr increase of natural gas, which is approximately 3% of the feed acid gas, resulted in an 8% increase in combustion air demand (or air flow requirement). This is equivalent to a decline in SRU sulfur inlet capacity of approximately 6%. For this case study, the required temperature in the reaction furnace was achieved by introducing natural gas at 8% of the feed acid gas—this caused an 18% reduction in SRU capacity. This value will vary depending on the actual feed composition, operating parameters and capacity. FIG. 13 shows the impact of the natural gas co-firing on SRU capacity and air demand.

The modeled TGTU amine system uses 40 wt% MDEA with a standalone regenerator. Although the optimized CO_2 can reach 85% slip, factors such as solvent over-circulation or the formation of primary and secondary amines from MDEA degradation will increase CO_2 pickup in the amine absorber. Therefore, the TGTU is modeled assuming a more conservative 70% CO_2 slip. The effect of the TGTU amine unit performance on the reaction furnace temperature is demonstrated in FIG. 14. As expected, increasing CO_2 slip correlates to less CO_2 in the combined AAG (and total feed gas) and a hotter reaction furnace.

The TGTU amine absorber inlet contained 70% more CO_2 in the co-processing case than it did in the crude oil-only case. This resulted in greater rich loadings and more H_2S leaving in the absorber vent gas. The rich loading could be reduced by increasing the amine circulation or by increasing the amine strength, or both. Increasing the circulation requires more energy consumption to circulate and regenerate the amine. Increasing the strength requires more amine inventory and additional energy to strip the amine. TABLE 7 shows the optimized case with increased amine concentration, which required 17% more steam to regenerate the amine. It should also be noted that, when operating with high amine concentrations, the amine losses will also increase. Therefore, when optimizing, operators should consider their requirements and available resources (such as pump capacity, amine inventory and steam capacity).

Takeaways. The impact of biofeed co-processing on downstream treatment units can be quantified with a well-designed test run, appropriate gas and liquid analytical capabilities, and reliable process simulation. Based on this, a fit-for-purpose treating scheme that minimizes capital and operational expenditures and prevents unanticipated reliability issues can be designed and implemented.

A major component in the changes associated with biofeed co-processing is the additional water produced. This will usually require modifications or increases to the SWS system. If these are intelligently implemented, refinery freshwater requirements can often be significantly reduced. The additional water volume increases the unit's stripping steam requirement and will likely reduce the residence time of the water-liquid hydrocarbon separation vessels, which potentially will leave more liquid hydrocarbon in the wastewater. The total hydrocarbon content—or worse, a fluctuating content—

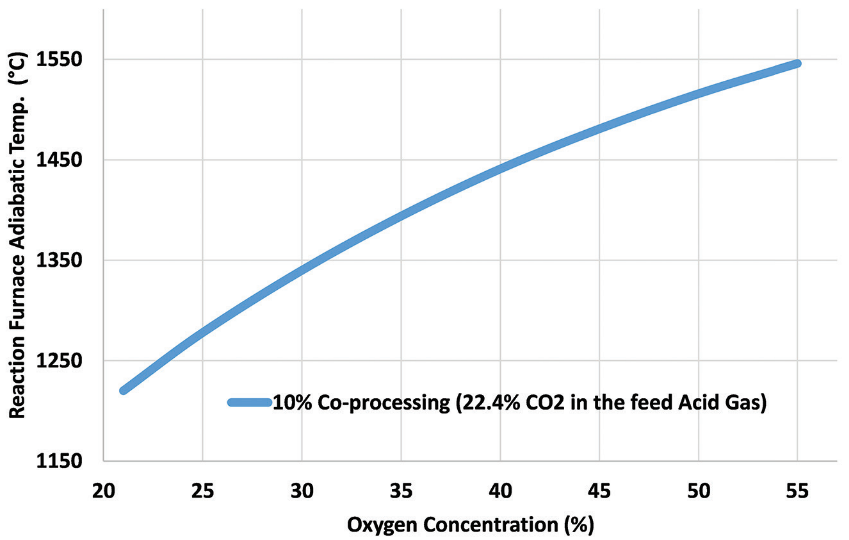

FIG. 12. Oxygen enrichment concentrations vs. front-zone reaction furnace temperature.

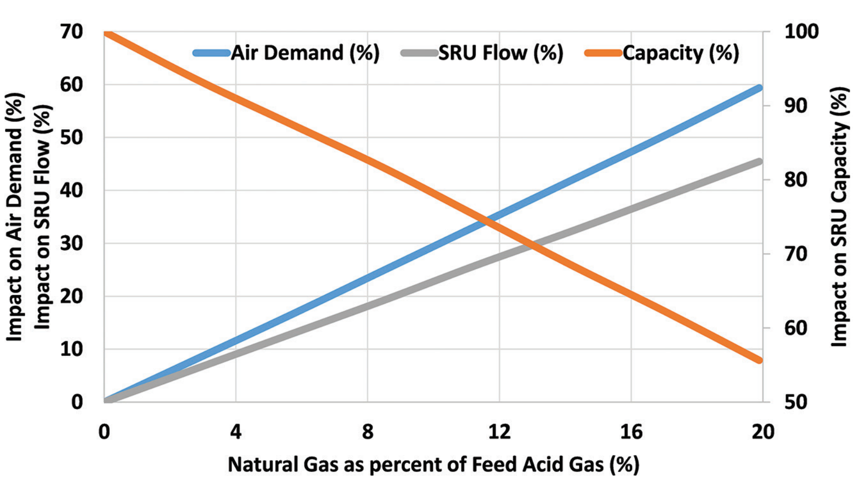

FIG. 13. Natural gas co-firing vs. the impact on air demand and SRU capacity.

can have detrimental effects and must either be minimized or mitigated.

Another major process change associated with biofeed co-processing is an increase in total system CO_2. The main amine system is generally able to handle the additional CO_2, although an amine system already at its limits may require additional (usually heat exchanger) equipment installation to manage the extra CO_2 picked up by the amine solvent. In some instances, changing the solvent can provide the additional capacity needed without capital investment, and can offset the impact of additional CO_2 on the downstream SRU by slipping more CO_2 into the product hydrocarbon streams. Depending on the co-processing methodology, foaming and emulsification issues can be severe, and normal management techniques such as anti-foam injection and carbon bed filtering may be insufficient.

The increased CO_2 concentration in the combined sour feed gas to the SRU should not be a problem for already-rich feeds (> 90% H_2S). However, for leaner feeds, care should be taken to ensure that the reaction furnace temperature remains at a minimum of 1,250°C to avoid contaminant breakthroughs. This can be done by either co-firing with a fuel gas of stable composition, installing an oxygen enrichment system where burner turndown capability allows, or by increasing the bypass fraction in the case of a split-flow furnace design.

As was the case with the main amine system, the absorber of an amine-based TGTU will encounter more CO_2. Amine circulation rates, amine strength and solvent selection can be adjusted to increase the CO_2 slip in the absorber, thereby preventing it from diluting the SRU feed stream, while holding the H_2S concentration in the absorber vent gas to specification. Nonetheless, the extra CO_2 will also require more duty from the TGTU regenerator reboiler and reflux condenser.

Understanding how biofeedstocks interplay with existing refinery infrastructure is critical for achieving smooth, easy and cost-effective co-processing. The effects and solutions listed in this article should serve as a guideline for determining the key parameters to monitor at a facility, along with ways to minimize operational issues as the world moves toward more green fuel sources. HP

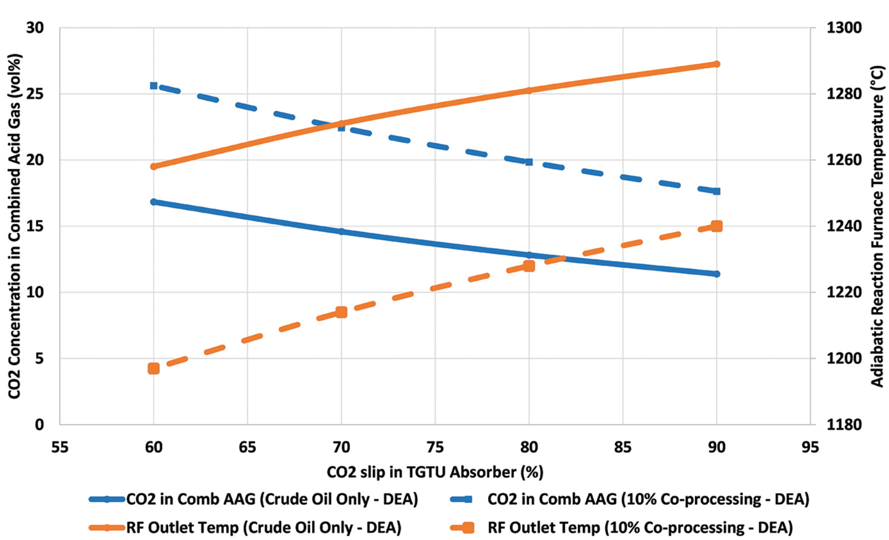

FIG. 14. The effect of CO_2 slip on reaction furnace temperature and inlet CO_2 concentration.

NOTES

[a] Optimized Gas Treating's ProTreat™
[b] Aspen Technology's Aspen-HYSYS™

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Productivity in process plant engineering: Aspects and examples

For centuries, division of labor has been the recipe to improve productivity—dividing an organizational set of tasks into smaller tasksets, as assigned to individuals.

In his guidelines, Fayol[1] separates a set of tasks to execute the core function from sets for support (i.e., line and staff); he also divides decision-making. Taylor[2] focuses on splitting the line functions into small tasksets that would be further adjusted for—apart from discipline—experience and grade, senior/junior, among other criteria that are easily picked up, enhancing productivity. Decades later, it was shown that very narrow tasksets reduced productivity, an approach that was later amended; consequently, some decisions were re-added. Another change is a reduced degree of freedom to divide into tasksets; people are prepared by education and by expectations. Still, the fact still stands that smaller tasksets are easier to learn. Choices are to be balanced for educated skills and experience, and for creating lateral decision-making—thus enabling personal development, which pays off in productivity.

These basic divisions focus on executing a task, with a limited distinction of the core treats in each taskset. This is less relevant for manual and rule-based tasksets but are more relevant for design tasksets. The underlying design process is to create solutions that can be transformed into instructions for fabrication. These solutions depend—in varying degrees—on creativity, and on tasks such as calculation, drawing and programming. Creativity requires iteration. Architects, and aerospace and process plant designers all iterate to create solutions. Even in detailed design, there is iteration: to correct, to improve quality, and to align to norms and fabrication. If a designer works on a repeat, then the need for iteration is limited and engineering is 5% of capital expenditure costs. If innovation is required, then the rate of iteration increases. The productivity question now includes how to iterate as efficiently as possible.

Iteration and productivity. Iteration stands for "redoing" with new inclinations: to correct, validate, adjust and learn by redoing. It can be the processing of ideas, not requiring any defined tasks. In practice, it is combining ideas and iterating by means of calculations and comparisons. Architects redo a concept, while process engineers redo a process design. An architect addresses space, and a process engineer addresses variables of a chemical process. The architect uses 3D tools, and the process engineer uses simulation tools. In design processes, the tasksets can be split into "create a concept" by one person and "redo" by another. Often, the division includes juniors to perform this iteration. Further into design, it may be a separation among processes and controls.

For productivity, this separation is to be reconsidered—for architecture, as for plant design—since separate tasksets can, supported by IT, be merged again. Results of automated checks and digital plants provide feedback on initial concepts. This feedback reduces time spent on iteration (up to more than tenfold). In addition, 3D tools allow users to iterate in less than 20% of the traditional time. Integration of process simulation with piping and instrumentation diagrams (P&IDs) can increase the productivity of iteration tenfold. Supported by IT, iteration increases productivity. If 50% of design is typically spent on iteration, this rate can be reduced to less than 10%. This is a major jump in productivity.

The extra price tag of iteration in plant design can be 100% vs. repeat projects. Therefore, focusing on iteration to enhance productivity is—in non-repeats—as much a driver as dividing tasksets: working on task structures and tools that reduce iteration costs are as critical as the division into these. To explore such options is to assess where the iteration is to take place and where to select tasksets and tools to reduce iteration cost.

This must be considered if the organization is working on innovative designs, or if lead time is relevant or if quality issues are at play. For lead time, iteration needs to be confined to the first steps in the process. For quality, the focus relates to improving specs. For innovation, it is essential for exploring concepts. For cost, iteration is to be avoided.

Analysis of the design process to improve productivity, including iteration. The actual functioning of the process can be analyzed to assess potential improvements. From initial

design (e.g., front-end engineering design) through the preparation of the ISOs, the design process tasks can be detailed. The next challenge is to find how the human tasksets and tools are mapped to the tasks. For example, in a refinery, the design is split according to scopes (scope may be set for other reasons, such as set boundaries to control development), each with its own assigned process (for capturing hydrocarbon or to use intermediate output as energy, each with a need to innovate).

To improve the productivity on iteration, one first assesses the sequence of tasks and depicts the type. In front-end design, this takes place around exploring options and optimizing choices based on process efficiency and capital expenditure costs. This concentric process is less predictable and differs from one that is triggered by a client who considers a value change in the original scope. This may be seen as disturbing, even if it creates a value-added process. Iterations necessary to correct design products are seen as a loss. Their actual costs can be low if added early in the process but can be substantial when added late in the process.

Higher productivity is a function of the tasksets and of project coordination. If iteration is assigned to a separate taskset, it requires more coordination. Enabled with IT, iteration may possibly be included in the taskset from which it was split. Similarly, faster CAE systems enable iteration in existing tasksets. If the goal is cost, one must limit this iteration, since it is a productivity leak. If the goal is innovation, integrated tasksets can iterate at a low cost. For quality, one can still consider assigning iteration to juniors to iterate efficiently.

The iteration levels in the design process can be measured during each project execution. For example, change orders and corrections are logged and indicate when iterations are triggered. It is timing and the tools that are used. The timing of process simulation or of sharing equipment specs and P&IDs illustrates iteration, as well. Over time, the pattern can be visualized.

The next step is to match the actual pattern with the projected pattern, related to the project goal. Projects to be considered are those of at least 20 yr. If the goal is innovation, iteration should be facilitated for a lengthy period, and possibly pushed up further when issues occur. If the goal is saving on lead time, iteration is to be as early as possible. If the goal is to save on costs, then iteration is to be avoided—corrections are to be made early in the process, as delay will eat away profits.

The necessary iteration patterns are now to fit the tools that are used in tasksets. If the tools for concentric iteration are fast, productivity can increase by merging previously divided tasksets. If a design is very complex, there is a need to be careful to tinker with the tasksets, but it may improve with iterative support from juniors. If a design is less complicated or requires quality iteration, tasksets can be brought together again (e.g., merge piping with structural, civil and control criteria). If the project has several areas, one can assign new tasksets for recurring units, such as for units for cooling gas in LNG plants. In all cases, productivity has a division plus an iteration dimension.

Choices in tasksets and related iteration. Dividing in tasksets is a valid approach to improve the productivity in less-iterative tasks and to perform these at a lower cost per hour. It implies extra coordination cost that must be factored in. When iteration is needed, and when it touches more tasksets, the cost of coordination increases, and datasets must be updated in various domains. Therefore, multiple tasksets inhibit iteration. If the iteration is substantial, division in tasksets increases iteration costs. Adding up both costs, a reduced division—supported by IT, which integrates geometric and technical data—makes the iteration more efficient and surpasses the efficiency obtained by the division in tasksets.

This tradeoff pays when iteration is expected in areas where innovation is needed, or in confined areas where quality is needed. Such steps are least effective when engineering costs are low. Lead time limits iteration to a confined first short period.

Therefore, where the focus is on quality and innovation, one benefits from integrated IT and a reduced division in tasksets. When working on productivity in cost-oriented engineering, iteration is to be confined and costs of integrated IT may not pay off. If organizations face different goals, then the goal of each project defines productivity possibilities.

Examples of division of tasksets and optimizing iteration toward the project goal. Examples may clarify the steps to increase productivity. In many projects, there are recurring units whereby the right IT can fuse tasksets, limiting coordination while supporting iteration (TABLE 1). There are units where the engineering product can be a computer numerical controlled (CNC) program for fabrication, replacing the drawing that must be prepared by the fabricator at a cost. In energy, carbon dioxide (CO_2) reduction is an integral part of revamping refineries. One must apply innovation on a unit in a project that is cost driven.

Example 1: Pump and compressor units. Units for pumping and pressurized media recur in many plants. These units are usually engineered by the traditional division in the design process: mechanical, controls, procurement, piping and civil. None of these tasksets must work very iteratively, but they require coordination. With integrated and dedicated IT for many of the units, the tasksets could be merged into one set, thereby saving on coordination costs and reducing engineering unit costs. One IT-supported engineer who performs all disciplinary tasks can save at least 25% of engineering costs on such units.

Example 2: LNG cooling structures. LNG plants are developed around the capacity of compressors and cryogenic heat exchangers. If the trains are limited and require a smaller gas-cooling structure, they become standards procured from pro-

TABLE 1. Examples of optimizing iteration towards project goals

Examples/aspects	Goal	Core direction productivity	Division of labor	Iteration	Savings unit
Pump and compressor unit	Engineering costs	Coordinate and execute low cost	Fuse disciplinary tasks	Limit cost iteration	More than 25% engineering
Gas cooling unit LNG	Fabrication costs	Integrate disciplines, prep CNC	Fuse tasks and include prep fabrication	Automate iteration	15% CAPEX
CO_2 capture unit	Innovation	Facilitate iteration	None	Lower iteration costs	20%–30% engineering

viders. For larger units (e.g., capacities of more than 4 MMtpy), the cooling structures on each project are designed. Here, it pays to create a new integrated engineering taskset with integrated IT support to cover structure, civil, piping and controls tasks. By doing so, the efficiency of engineering the unit is improved and CNC programs for fabrication replace ISOs. Total capital expenditure savings in such units may be 15%–20%.

Example 3: Innovation for CO_2 recovery units or for alternative low-CO_2 feedstock. Investments in refineries to reduce their CO_2 footprint require innovative engineering. Option 1 is to capture CO_2 when processing hydrocarbon (i.e., carbon capture and storage). Cost-effective solutions are needed, pursued by confined iteration to improve the quality of a solution. For these functions, one pursues a lower division of labor among highly qualified engineers, supported by integrated IT. This may be expensive, but it will lower the iteration cost. The division can include juniors to iterate. On quality, effective IT to iterate is needed. This innovation and quality drive still fits divided tasksets. Savings in the innovative scope are 20%–30%.

Option 2 is to work on new forms of feed. This requires more innovation. It may even require a low division of tasksets and integrated IT to ensure results. These steps are safeguards to produce tangible results. Avoiding a failure to innovate adds to savings.

Takeaway. These examples show that—in a greater drive for improvement—substantial results can be achieved within engineering. Division of labor is still instrumental for productivity, although it now needs a different approach in which iteration is treated as a separate feature, while IT is instrumental for productivity gains. This assists to tune design organizations to improve their competitive standing. Two examples illustrate cost savings through fusing tasksets and one through integration of design and fabrication data.

Therefore, there are various productivity-enhancing solutions that push the design organization to more productivity. These can lead to healthy profit margins, based on a project level guidance toward innovation and quality performance and the ability to perform projects that require low design costs or short lead times. It is a challenge for engineering, procurement and construction (EPC) companies to increase their current 1%–3% margins to 6%–7%. The combined steps listed in this article certainly elevate margins on projects. HP

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Piping interface type selection

Most oil and gas process plant projects involve multiple work packages executed by different lump sum turnkey (LSTK) contractors executing work within their own package, with a final piping interface connection at an agreed boundary limit to accomplish the interface joint between two parties and successfully close the piping interface.

An interface point (IP) is a physical location or boundary where responsibility for the item passes from one contractor to another contractor. Piping IPs are commonly located at or near a geographical contract boundary, generally known as an interface gate (IG). One of the contractors is assigned the role of lead contractor, which has the responsibility for the delivery of the IP for final closure while also determining the interface type based on any limitations. The counter party to the lead contractor is known as an interface contractor, which is responsible to furnish necessary material and support details up to the agreed IP.

For LSTK projects that do not have a predetermined set of criteria for IP type selection, this article aims to establish the best selection of above-ground piping interface type at the contractor package boundary based on the following parameters:

- Advantages and disadvantages of different types, with consideration to any effects on the integrity of the interface point
- Impact on construction cost and schedule
- Seamless transfer of responsibility for a physical interface connection from one party to another, and plant reliability, availability and maintainability (RAM) during its design life.

For LSTK projects, it is recommended to review relevant project specifications and guidelines that show any technical limitations for interface types. Two interface types—welded interface and flanged interface—are further evaluated for above-ground piping interfaces between two parties for any process or utility service.

Other piping interface types, such as proprietary mechanical piping joints for metallic pipes, and bell and spigot, lamination or butt-fusion joints used in nonmetallic pipes, are outside the scope of this article.

INTERFACE TYPES

A brief overview of two interface types is provided here.

Welded interface. An interface connection between two contractors at a predefined IP location is achieved via a welded joint between two pipes, as depicted in **FIG. 1.**

An appointed lead contractor will be responsible for the final closure weld and for testing in accordance with test procedures. If remedial work is necessary, the appointed lead contractor normally conducts this work as specified in the project contract.

A welded interface refers to a butt-welded joint between two pipes following correct pipe end preparation by an individual contractor within its own scope to make a proper final weld joint. Contractors must exchange, agree on and finalize the relevant details of the piping interface before executing the interface work, such as:

- Pipe—pipe size, material, wall thickness, insulation, end preparation, etc.
- Plant coordinates—Easting, Northing and Elevation to specify exact location in 3D space within the IG.
- Stress and support—stress analysis results to confirm allowable stress limits, support types and their locations leading up to the IP.
- Fluid—fluid name or service and other properties such as fluid phase, fluid temperature, pressure, flowrate, flow regime, velocity, viscosity and all such relevant operating, design and any upset conditions of the line in

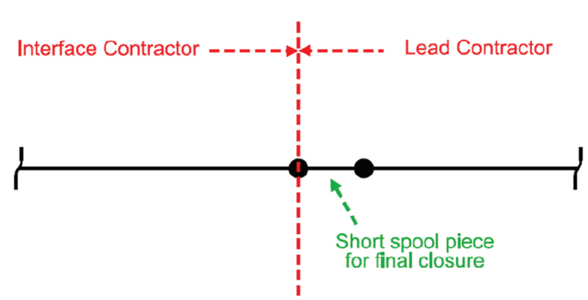

FIG. 1. Welded interface.

any operating scenario throughout the life of the plant, typically noted in the line designation list.

Flanged interface. An interface connection between two contractors at the predefined IP location is achieved via a flanged joint between two pipes, as depicted in **FIG. 2.** There may or may not be a valve at the interface point, as agreed between both parties.

The appointed lead contractor will be responsible for the supply and installation of gaskets and bolts and for the final system test before hand-over activity, as specified in the project contract.

The flanged IP refers to the individual contractor supplying the pipe spool with a flange at the end of pipe, or a valve with a flanged end if a valve is required. Similar to the welded interface point previously mentioned, both contractors must exchange, agree on and finalize the relevant piping and process related interface information before executing the work. Additionally, for the flanged interface point, both contractors must agree on the flange type, flange full description, flange facing and finish, gasket and bolts to ensure both flanges within the individual contractor boundary are compatible to make the final connection. If a valve is required at the IP, then all relevant information must be agreed beforehand. The flange face will generally be a flat face, raised face or a ring-type joint, as per relevant project specifications and guidelines.

Review of limitations. Limitations observed from project specifications, guidelines and American Society of Mechanical Engineers (ASME) B31.3 code[1] are summarized here.

Welded interface. Generally, project standards and specifications may not have limitations for welded piping joints, other than to comply with various requirements stipulated in ASME B31.3 and other such standards, including the use of approved welding procedures. When a flanged connection is preferred over a welded connection, exceptions include:

- When a dissimilar metal joint is involved to avoid galvanic corrosion or wet hydrogen sulfide (H_2S) damage
- A metallic-to-nonmetallic piping connection
- An underground to above-ground piping connection.

Flanged interface. Generally, project standards and specifications may show the following limitations for flanged joints:

- The limitations of using different flange types (e.g., slip-on, lap joint and compact flanges).
- The use of flanged connections may be prohibited when butt-welded joints can be used (e.g., cyclic or vibration services, potentially toxic material, steam of ≥ 900# pressure rating, piping subject to large bending or external loads)
- When project specifications recommend minimizing flanged connections in internally fusion bond epoxy (FBE)-coated piping systems
- Limitations of the allowable misalignment of flanged joints/bolt holes
- Project specifications, guidelines and ASME B31.3 provide guidance for pipe spool fabrication/erection about bolt straddling, offset of flange bolt-holes and permitted misalignment of flanged connections
- Project specifications provide bolt-tightening requirements.

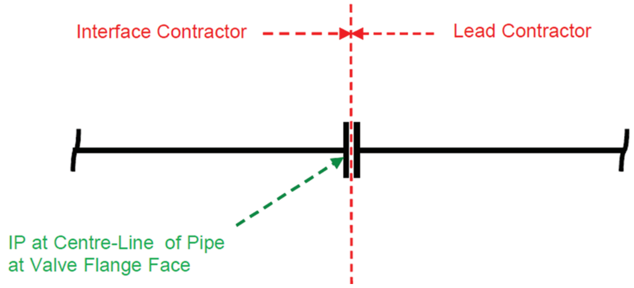

FIG. 2. Flanged interface.

WELDED PIPING INTERFACES

Advantages. Welded joints are considered technically superior to flanged joints due to better joint integrity and/or leak tightness. With welded joints, no special precautions and limitations are required as applicable to flanged connections, such as flange misalignment, concerns with overstressing or distortion of the bolts, correct selection of gaskets and bolts, etc.

Welded joints also have an advantage over flanged joints since after the welding and any subsequent post weld heat treatment (stress relief), as required, the piping system is considered one continuous homogeneous system. This not only offers improved thermal integrity of a joint when exposed to an external source of heating or fire, it offers improved stress distribution for a piping system under internal pressure and expansion or contraction, as well as under different temperature conditions the pipe may endure during its design life.

Additionally, welded piping interfaces are considered more beneficial from cost and schedule perspectives for the following reasons:

- Capital cost savings due to the use of two flanges and associated gasket and bolts saved per piping IP. For larger pipe sizes and ratings, the savings can be substantial. This also reduces the dead weight of the piping system.
- Since the majority of the piping components within the piping system (pipes, fittings, flanges, etc.) are joined via butt-welds, adding an extra butt-welded joint executed at the worksite at the IP limit can be considered an incremental cost only where it will replace the welded joint between the pipe and flange used for a flanged interface.
- A savings in engineering work hours to analyze the flanged joints for the leakage checks, as required by project specifications. Since this check is conducted at IPs, it will require further exchange and agreement of information between two LSTK contractors before the execution of an IP.

Disadvantages. The main disadvantage of welded joints in the context of this discussion is that they do not offer electrical discontinuity desired for certain cases, such as welding between two dissimilar piping materials. A welded piping interface with two dissimilar materials, such as carbon steel and stainless steel

pipe, will cause galvanic corrosion, making carbon steel more vulnerable to corrosion. Furthermore, dissimilar welds are generally not allowed in wet H_2S service. For such cases, welded interface is not recommended and flanged IPs should be used.

Another disadvantage is the hydrotesting of the interface weld joint, which requires special tools and isolation for items that have already been hydrotested.

FLANGED PIPING INTERFACES

Advantages. Flanged joints offer reduced construction lead time onsite due to prefabricated pipe spools with flanged ends at the IP that only require time for the mechanical fitment, while avoiding the hot work (welding) and other quality control processes such as post weld heat treatment and weld inspection required for the welded interface points. In certain conditions, executing a field weld for a welded IP will be more time consuming and may involve a higher cost if the majority of the welds are conducted at a shop with no or a limited number of field welds at a given IG location.

Flanged joints also offer an opportunity to use two dissimilar piping materials, allowing an individual LSTK contractor to use the piping material to suit its package design conditions, and a more economical design that uses insulating gasket kits at the flanged IP to avoid any concerns related to galvanic corrosion.

Similar to a welded IP, flanged joints can also be inspected in a non-intrusive manner for corrosion on the flange faces by existing techniques, such as phased array ultrasonic testing, without needing to dismantle the flanged joint [as per American Petroleum Institute (API) RP 574].

Disadvantages. The main disadvantage of a flanged joint at a piping interface is potential leaks, which could have profound consequences including safety hazards that can affect plant availability.

Further, a flanged IP does not reduce the number of welds in a given piping system. Each LSTK contractor must provide a flanged connection where a flange is welded to the pipe within its own boundary with the final gasket and bolts provided by the lead contractor. For a welded IP, only one final weld joint by the lead contractor is required; however, for a flanged IP, each contractor must provide a pipe-to-flange welded joint in its scope.

Flanged joints will occupy more space and will require the staggering of flanges if two adjacent pipes are both flanged at interface points. Additionally, the bolts and nuts on flanges may require inspection at periodic intervals to ensure the integrity of the flanged joints and for potential corrosion on bolts, particularly when these flanged joints are under insulation. Each time a flanged assembly is opened, the gasket must be replaced as per project specifications and guidelines.

The majority of piping IPs are generally located in non-operating/interconnecting areas of the plant typically outside the plant battery limit and infrequently accessed by plant operations. This requires further inspections or walk-throughs to ensure flanged joint integrity (e.g., flange loosening or bolt distortion due to pipe movement, corrosion on bolts, loosening of any insulation). Where flanged joints are insulated, it will be more expensive to insulate a flanged IP compared to a welded IP due to greater surface areas to be covered by the insulation.

Takeaway. The individual comparison between the two types of piping interfaces shows that welded interfaces are preferred and should be prioritized over flanged interfaces. Flanged IPs should be used only where welded interfaces are not desired, such as in the case of two dissimilar or incompatible piping materials at the IP between two LSTK contractors. HP

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The 1930s: Catalytic cracking, polyethylene, synthetic fibers, resins and jet engines

The hydrocarbon processing industry (HPI) has a rich history of discovery, challenges, breakthroughs, trial and error, collaboration and success. *Hydrocarbon Processing* continues its reflection on the history of the HPI. In the January issue, a detailed analysis was provided on the origins of the modern refining and petrochemicals industries. This included the discovery of kerosene, the construction of new refineries around the world, the production of the first synthetic plastics, the rise of the internal combustion engine (ICE), oil demand's exponential growth during and after World War I (WWI) and how thermal cracking evolved refining processing. The following will detail how the HPI continued to evolve during the 1930s.

The discovery of catalytic cracking. After serving in WWI in the French artillery division and later in the tank corps, French engineer Eugene Houdry worked in his father's steel business. Outside of work, Houdry had an interesting hobby, racing cars. Through these endeavors, he began to develop a passion for improving engine performance.

With the significant increase in gasoline demand after WWI, many forecasters feared that thermal cracking was unable to satisfy future global demand. Like many other researchers around the world, Houdry was trying to develop a new way to develop high-performance fuels.

In the late 1920s, he and French scientist E. A. Prudhomme developed a three-step process to convert lignite to gasoline. However, a major problem with the process was that the catalysts would get coated with carbon, lessening their effectiveness. To solve this challenge, Houdry used Fuller's earth (a naturally occurring aluminosilicate), which effectively produced gasoline from lignite.[24] However, pilot plant demonstrations yielded less than expected results and the process was deemed uneconomical.

Unable to secure additional financial backing, Houdry moved to the U.S. and eventually started working with Socony-Vacuum and Sun Oil Co. to perfect his process. In 1936, the first Houdry Unit began commercial operations at Sun Oil's Marcus Hook refinery in Pennsylvania—it was the first fixed-bed catalytic cracking unit.[23] Approximately 50% of the 15,000-bpd unit produced high-octane gasoline, which was double the production of conventional thermal processes.[24] Around the same timeframe (circa 1938), the alkylation process was commercialized in the U.S. The process produced high-octane aviation gasoline, which saw significant demand increase during World War 2 (WW2). The process was then used in the 1950s to produce blending components for automotive fuel.

The catalytic cracking process was later improved upon by Warren Lewis and Edwin Gilliland while working for Standard Oil of New Jersey (U.S.). According to literature,[25] the improved process included a continuously circulating fluidized catalyst made of fine zeolite powder. Houdry's fixed-bed unit gave rise to research and development by other companies that led to the invention of the fluid catalytic cracking process in the 1940s.

Coking and gasification evolve. The first delayed coker was built in 1929 by Standard Oil of Indiana (the company would later become bp). The Burton thermal cracking process produced coke that would be sent to a vertical coke drum. However, cleaning the vertical coke drum required arduous manual labor. It was not until the late 1930s that Shell introduced hydraulic decoking at its refinery in Wood River, Illinois (U.S.), which used high-pressure water to clean coke drums. This process enabled refineries to use two coke drums for continuous operation.[26] Over the next several decades, coking would become a staple in refining operations.

In the mid-1930s, Lurgi GmbH (now a part of Air Liquide) invented a novel coal gasification process. The pressurized, dry-ash, fixed-bed gasifier would use coal to produce synthesis gas (syngas). The first commercial Lurgi dry-ash gasification plant started operations in 1936, and the process is still in use today.

Polyethylene: An accidental discovery. In 1933, while working at Imperial Chemical Industries (ICI) in Northwich, England, Eric Fawcett and Reginald Gibson stumbled upon a white, waxy substance during experiments they were conducting on ethylene and benzaldehyde. The experiments included heating the mixture to 170°C at an extremely high pressure—more than 1,900 bar—in an autoclave. However, the reaction was a safety hazard due to the explosive nature and research was halted.

Two years later, ICI scientists Michael Perrin, John Paton and Edmond Williams began to conduct additional research on Fawcett and Gibson's discovery. In this iteration, the scientists repeated Fawcett and Gibson's test but focused solely on ethylene. What the three did not know was that the pressure vessel used leaked, leading to a loss of pressure. Once the reaction was completed, the trio noticed a white powdery substance remained—one lab technician described the substance looked like a lump of sugar.[27] The scientists had accidentally stumbled upon polyethylene, which would revolutionize society.

Over the next several years, ICI perfected the process and found practical uses for the material (the first item ever made with PE was a walking stick[28]) that would not only produce products to modernize society but also aide the Allies in WW2.

ICI produced the first ton of PE in 1938 (FIG. 1). In 1939, the first commercial-scale PE plant went into operation. The 100,000-tpy plant was instrumental in producing PE on an industrial scale. Within the next few years, many PE plants went into operation, primarily to aide in the allied war effort. PE was used extensively as insulating material for radar cables during WW2. The material was lightweight, which enabled Britain to install radar in their fighter planes, providing a significant technical advantage in long-distance air warfare.[27,28] Due to this wartime advantage, the production of PE for insulated cabling was highly-secretive. It was not until post WW2 that the production of PE was commercialized. Within several years, PE production capacity significantly increased and would later become the world's most used thermoplastic.

FIG. 1. A commemorative sample of the first ton of PE produced by ICI in 1938. The initials G. F. are that of George Feachem, a chemist that was on duty the night PE was produced in laboratory tests in 1933. According to family members, Mr. Feachem kept this token in his wallet until his death. Photo courtesy of BBC History of the World.

FIG. 2. Carothers demonstrates the elasticity of neoprene. Photo courtesy of the Science History Institute.

New chemical discoveries with lasting legacies.

Several new chemical discoveries took place in the 1930s that have provided the global population with new products to improve standards of living. These included the discovery and production of polystyrene, polyepoxide, nylon, polyester and neoprene.

Polystyrene. Although discovered in the late 1830s, styrene—which would lead to the production of polystyrene—would not be commercialized for nearly 100 yr. In 1839, German chemist/pharmacist (referred to as an apothecary) Eduard Simon distilled an oily substance from storax, a resin from a sweetgum tree. He noticed several days later that the material—which he called styrol—thickened into a jelly-like substance. Thinking the reaction was due to oxidation, Simon termed the substance styrol oxide.[29] However, it was not until 80 yr later that a practical use was found for the material.

In the 1920s, research/writings by German chemist Hermann Staudinger led to the invention of polystyrene. Staudinger demonstrated that thermally processing styrol produces macromolecules, which he characterized as polymers. His technical research/writings would eventually lead Staudinger to be awarded the Nobel Prize for Chemistry in 1953.

Commercialization of styrene polymers began in the early- to mid-1930s by IG Farben in Germany and Dow Chemical in the U.S.—styrene production increased significantly in both Germany and in the U.S. during WW2 to produce synthetic rubbers to aide in the war effort. In the late 1930s, Dow Chemical engineer Ray McIntire was experimenting with a polystyrene process developed by Swedish inventor Carl Munters. By accident, McIntire created foam polystyrene which expanded approximately 40 times in size.[30] Dow would later commercialize this discovery as expanded polystyrene, better known and marketed under the name Styrofoam.

Neoprene and nylon. While focusing on research conducted by Staudinger and Belgian-born priest and chemistry professor Julius Nieuwland, Wallace Carothers' polymers research group at DuPont discovered two major chemical applications: neoprene and nylon. While a professor of chemistry at the University of Notre Dame (U.S.), Nieuwland focused his research on acetylene chemistry, which led to the discovery of divinyl acetylene—a jelly-like substance that hardens into an elastic compound similar to rubber.[31] DuPont purchased the patent rights to this new discovery, and Nieuwland joined Carothers' research team to conduct further research and testing on practical applications for this and other polymer applications. One of Carothers' colleagues, Arnold Collins, discovered neoprene while conducting further research on divinyl acetylene. Through several testing methods, Collins soon discovered a mixture that produced a clear homogeneous mass that bounced. The product—chloroprene—was used to produce the polymer polychloroprene, which later became the new synthetic rubber neoprene.

Neoprene was first marketed in 1931 under the name DuPrene; however, the product was re-envisioned since it contained an odor due to the manufacturing process. By the mid- to late-1930s, the improved product—suitable for many applications (construction, automotive, medical equipment, fabrics, electrical equipment, textiles, among others) and marketed under the generic name neoprene—generated substantial revenues for DuPont.

After the discovery of neoprene, Carothers' team turned their sights on producing synthetic fibers. By the mid-1930s, Carothers produced fibers comprised of amine, hexamethylene diamine and adipic acid. However, water produced during the condensation reaction process would fall back into the mixture, preventing the creation of more polymers. After adjusting the process, Carothers pro-

duced strong, elastic fibers.[32] The new material was called polymer 6,6 (or nylon 66) since the two monomers that comprise the substance each contained six carbon atoms (FIG. 2). Nylon first became a household product as women's hosiery, later being used in the U.S. war effort to produce parachutes and tents. Over the next several decades, nylon would be used extensively as a combined fabric in fashion and apparel, as well as in several industrial applications—the global nylon industry market size is forecast to reach more than $46 B by the late 2020s.[33]

Polyester. Carothers' research also led to the discovery of polyester in the early 1930s. However, the discovery of nylon pushed additional research on polyester to the backburner. It was not until the late 1930s that British scientists John Winfield and James Dickson expanded on Carothers' work on synthetic fibers. Their research would eventually lead to the development of polyethylene terephthalate (PET) in 1941, which they marketed under the name Terylene. DuPont would later purchase the rights of the British scientists' discovery and develop a new synthetic fiber in the mid-1940s they called Dacron. In the early 1970s, PET began to be used in the production of plastic bottles, and today, PET is the fourth most produced polymer after PE, polypropylene and polyvinyl chloride.

Resins, epoxies, polyurethane and Plexiglas. The DuPont company was not finished with major polymer discoveries of the 1930s. In 1938, Roy Plunkett was assigned to research chlorofluorocarbon refrigerants to find a better way to refrigerate food. Much like the discovery of PE, an accident led to the discovery of another important product still in use today. Plunkett stored 100 lb of tetrafluoroethylene gas in small cylinders at dry-ice temperatures [approximately –78°C (–109°F)] before chlorinating it. When he opened the cylinder, instead of gas pouring out, Plunkett noticed a white powder had formed (FIG. 3).[34] Further investigation found the substance to be heat resistant and had a low surface friction. DuPont polymer scientists determined that the tetrafluoroethylene gas polymerized to produce the material, which DuPont would later market under the name Teflon.

In 1936, while working at Monsanto Chemical Co., William Talbot produced melamine formaldehyde by polymerizing formaldehyde with melamine. This new substance was a thermosetting plastic that was very good at maintaining strength and shape. Melamine resins were used for many different applications, including in utensils, plates, furniture, cups, bowls, laminates, toilet seats, automotive and epoxy coatings, among others.[35]

Within the next 3 yr, other significant chemical discoveries were made. In 1936, British chemists John Crawford and Rowland Hill discovered polymethyl methacrylate (PMMA) while working at ICI in England. PMMA is a clear thermoplastic resin that is more transparent than glass and 6x–7x more resistant to breakage than glass.[36]

Around the same time, German chemist Otto Röhm conducted experiments with methyl methacrylate (MMA). One experiment involved polymerizing MMA between two layers of glass in a water quench. The result was a clear plastic sheet that was lighter than glass but much less prone to shatter. Röhm's chemical company—Röhm and Hass AG—soon marketed the material under the name Plexiglas (FIG. 4). According to literature, the first major applications of the new plastic were for aircraft windows and bubble canopies for gun turrets during WW2.[37] After this discovery, several companies around the world developed their own PMMA products under various proprietary names. Röhm and Hass AG's business lines were eventually acquired by different multinational businesses, including Dow Chemical, Arkema and Evonik.

In 1936, while working with synthetic resins to produce dental prosthesis, Swiss chemist Pierre Castan developed a solid by reacting bisphenol A with epichlorohydrin and curing it with phthalic anhydride. Castan's invention—epoxy resin—was first used for dental fixtures and casings,[38] later being licensed by Ciba Ltd., which would become one of the largest epoxy resin producers in the world.

FIG. 3. Plunkett (far right) and colleagues reenacted the discovery of Teflon in 1938. Photo courtesy of the Hagley Museum and Library.

FIG. 4. After discovering Plexiglas, Röhm and Hass AG marketed the material by saying, "The light and weather-resistant, practically unbreakable, flexible and easily formable Plexiglas is made from a new, viscous synthetic resin." Photo courtesy of Evonik Industries.

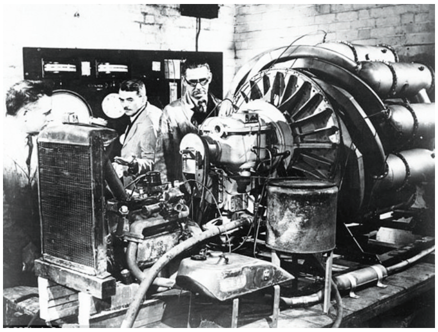

FIG. 5. Whittle and his colleagues work on the first jet engine. Photo courtesy of Getty Images.

Around the same timeframe, Sylvan Greenlee was conducting his own research on epoxy polymers in the U.S. by reacting epichlorohydrin with bisphenol A. His research created the epoxy resin bisphenol A diglycidyl ether (DGEBA or commonly abbreviated as BADGE), which would become the most widely used commercial-grade resin in the world. Epoxy resins are presently used in many industrial and commercial applications, including paints and coatings, adhesives, electrical systems and electronics, marine and aerospace applications, and many more.

While epoxy resins research and development were being implemented in Switzerland and the U.S., German chemist Otto Bayer was setting his sights on polymer research at IG Farben in Leverkusen, Germany. In 1937, Bayer created a new polymer by reacting 1,8 octane diisocyanate with 1,4 butanediol.[39] This new material was named polyurethane, which would later be used in many applications, including in construction, furniture, insulation, coatings, adhesives, sealants, elastomers, moldings, appliances, automotive, apparel and many more.

Ingenuity takes to the skies. Prior to designing engines, British engineer and inventor Frank Whittle was an airplane apprentice and pilot at the Royal Air Force (RAF) College Cranwell. Although garnering the reputation as a low-flying daredevil and aerobatics stuntman (not in a positive light), Whittle had an eye for airplane engine designs. In his graduation thesis *Future Developments in Aircraft Design,* Whittle believed that the evolution in flight would not be better propeller designs but the use of improved combustion engines for propulsion. In addition, he believed that airplanes would be able to fly faster (more than 500 mph) and farther at higher altitudes due to low air density.[40,41] However, when Whittle provided his concepts to the RAF, they were rejected as impracticable. Despite being rejected by his superiors, Whittle continued to publicize his jet engine concept and filed a patent for his engine design two years later in 1930. According to literature, the concept was a two-stage axial compressor feeding a single-sided centrifugal compressor, what he referred to as a "turbojet."[42] Whittle continued to work on building his jet engine designs over the next several years, forming Power Jets Ltd. in 1936 (**FIG. 5**).

Unbeknownst to Whittle, German physicist and engineer Hans von Ohain was developing a similar jet engine in Germany. Ohain joined aircraft industrialist Ernst Heinkel to design the Heinkel-Strahltriebwerk 1 (HeS 1) engine—German for Jet Engine 1. The first tests of HeS 1 were conducted in 1937. Although the tests were successful, the high-temperature burn—the engine ran off hydrogen fuel—scorched the metal, leading Heinkel and Ohain to switch to gasoline as fuel. Several changes were made to the design, and on August 27, 1939, test pilot Erich Warsitz flew a plane equipped with a HeS 3b centrifugal-flow turbojet engine—the latest iteration.[43] This historic day marked the world's first jet-powered aircraft flight.

Although Ohain beat Whittle to the first jet engine test flight, Whittle continued to improve his designs. As WW2 started, he received additional financial backing from the UK Air Ministry. In 1940, the first British jet-powered plane—the Gloster E.28/39—was flown using Whittle's W1A engine.[44] As war raged in Europe, the UK Air Ministry was ordering several thousand jet engines per month. By 1944, Whittle's engine design—produced by Rolls Royce—was used in the first British fighter planes, the Gloster Meteor, that could reach speeds of 600 mph.[45]

Over the next several years, jet engine designs continued to be optimized, primarily for military aircraft. However, on July 27, 1949, the world's first jet-propelled airliner made its test flight in England.[46] This historic occasion marked the first use of a jet-powered passenger plane, which would revolutionize travel. Over the next several decades, the jet-powered passenger plane would enable passengers to travel faster and farther in a shorter duration and build a nearly $200-B industry to carry billions of people each year to various destinations around the world.

The 1940s. As the world engages in conflict, demand for gasoline and chemical products soar to aid in the war effort. Post-WW2 will usher in new technological advances for producing higher octane fuels and chemical products that will increase the standard of living for hundreds of millions around the world. The industry's milestones of the 1940s will be discussed in the March issue of *Hydrocarbon Processing.* HP

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Operations, processes and safety evolve and advance: Excerpts from the 1930s

The following is a mixture of technical articles, columns and headlines published in the 1930s by *The Refiner and Natural Gasoline Manufacturer,* the forerunner to *Hydrocarbon Processing.* This collection of excerpts provides a look into the major technological advancements and topics/trends in the hydrocarbon processing industry during that timeframe.

API Division of Refining

January 1930

Through the realignment of American Petroleum Institute (API) activities to embrace three major departments (Production, Refining and Marketing), the organization has created the Division of Refining. This step is essential to improve operating efficiency and optimize the art of refining. How well the industry progresses collectively is governed by the willingness to contribute collectively.

The Division of Refining includes periodic regional conferences or conventions where tentative solutions of problems peculiar to a given refining district or, to the industry, may be discussed by refinery personnel in the various regions throughout any given year and may be the basis of highly interesting and instructive papers for presentation at group sessions at the annual API meeting.

Petroleum coke and its utilization

M. E. Schulz, January 1930

In some cracking plants, particularly those in which the heavier fractions of the crude are processed, a considerable amount of petroleum coke is formed. A refinery cracking several thousand barrels of oil per day may soon find itself with a large supply of coke on hand. What can you do with this product? We conducted some experimental work and found that coke could be used as a pulverized fuel. It is quite probable that this means of utilization will develop an important market for a large percentage of petroleum coke in some of the favorable localities. In the future, it is possible that this product will be converted from a material difficult to dispose of to a valuable byproduct of the oil refinery business.

Factors affecting the determination of gum in gasoline

C. R. Wagner and J. Hyman, January 1930

One of the more important problems that confronts the producers of cracked gasolines is that of gum formation. Such gasolines, if improperly refined, or if allowed to stand for long periods in contact with the atmosphere, seriously affect the operation of certain types of gasoline motors. Upon investigation, these motors show intake valves stuck tightly and the guides clogged with a black, brittle, carbonized deposit—referred to as the "gumming" effect of gasoline. The authors have been led to present the results of their work in the hope that they may be able to assist somewhat in clearing up the existing confusion among oil testing laboratories on this subject.

Survey shows need for curtailment of refinery runs

H. J. Struth, January 1930

A preliminary survey shows that refinery runs must be maintained at a rate not to exceed those of 1929. Forecasts show that global gasoline demand will increase 7.7% to 472 MMbbls (nearly 1.3 MMbpd).

Welding plays important part in refinery maintenance

January 1930

In present refining practice, pipe stills are replacing the older shell stills and the bubble tower or fractionating column has been adapted from the chemical industries to give clean cut separation of fractions in a single distillation. All this means that refinery equipment has suddenly become more complicated. Due to this trend, fabrication of intricate equipment is done right at the refinery, and welding has been of incalculable assistance.

Vacuum units lower costs and improve products

G. Reid, February 1930

Fire prevention and protection in oil refineries

F. L. Newcomb, February 1930

Fire prevention in oil refineries has four essentials, which are:

1. Design, which includes a clear conception of the use or occupancy of the property to obtain a safe layout with provision for future expansion, adequate strength for all structures and equipment, and the proper specification of suitable materials
2. Maintenance of structures and equipment, including systematic inspection to see that they are in a safe and fit condition for the service for which they are used
3. Care in all operations where flammable materials are handled
4. Good housekeeping or the maintenance of the entire plant in a neat, clean and orderly condition.

Securing maximum recovery from absorption

I. N. Beall, March 1930

This article is the first in a series written by Mr. Beall in the early 1930s focused on the practical application of technology to the manufacturer of natural gasoline. In discussing the problems confronting the manufacture of natural gasoline, he stated that he would like to discuss his several topics with the viewpoint that, "Theory is no better than its practical application proves it to be, and that theory without practice is as bad as practice without theory."

They are not accidents—They are self-imposed injuries

O. Peters, March 1930

Analysis of statistics leads to the conclusion that accidents are in fact self-imposed injuries. Records show that the first steps taken were in the provision of safe working conditions by the adoption of safety devices; then came the development of intelligent supervision; and now we are in the period of educating workers by teaching them the cause, the effect, the prevention and the responsibility of accidents.

Cutting losses by efficient vapor recovery

H. R. Linhoff, May 1930

Vapor recovery applies to the recovery and utilization of production, refinery and tank farm vapors, which if not recovered would otherwise be wasted. In the past, producers and refiners have been inclined to neglect the importance of vapor loss. These losses can be reduced in several ways (e.g. insulation, seals and floating roofs), but they cannot be eliminated without some system of collecting and recovering the vapors.

Operation of meters in natural gasoline plants

E. E. Stovall, May 1930

There are two methods of measuring gas that are used: the positive or displacement meter and the orifice meter measurement. This article focuses on reading the different types of meters, how to install them and computing measurements.

Progress in vapor phase cracking

C. R. Wagner, June 1930

The development of the high-compression gasoline engine has made it necessary for the refiner to market a fuel satisfactory for use in such engines. This trend has led to the commercial development of phase cracking in which the refiner is able to produce a highly anti-knock fuel from either topped crude or gasoil at a cost not out of line with the cost of gasoline made by liquid phase cracking processes.

Some theoretical considerations for the design of separators

F. L. Kallam and L. J. Coulthurst, July 1930

No two plants have the same separator design. It is self-evident that where practice is, so diversified theory is not understood. This fact has led to the present work of the subject of separators, with the hope that knowledge gained in other industries can be profitably employed toward correcting the separator failures in our own industry.

Methods of utilizing petroleum residuals and byproducts

A. E. Dunstan, July 1930

In discussions on the utilization of various petroleum residuals and byproducts for power purposes, the author focuses on important topics of acid sludge, petroleum coke, asphalt residues and waste gases. His work reviews various research, experiments and results obtained in the application of these products as fuels. Part 1 of this discussion focuses on acid sludge, while Part 2 covers the usage of petroleum coke, asphalt residues and waste gases.

Lime distributing plant to combat corrosion

R. J. Lawrence, August 1930

Considerable difficulty was experienced in refining certain crude due to corrosion on the crude tube stills and the battery of cracking units. On the cracking stills, this rapid corrosion soon gave rise to failure in tubes, return beds, headers and transfer lines. Immediately, several methods were implemented to reduce the corrosion rate, one being to employ hydrated lime. The lime treatment was chosen because enough of the alkali could be fed into the stills with the charging stock to greatly reduce the corrosion and still not be detrimental to the stills.

Few concerns dominate refining capacities

G. Reid, September 1930

Twenty-two companies own nearly 80% of the total crude capacity and 87% of the cracking facilities in the U.S. These 22 companies have 159 refineris and 1,713 cracking units. The remainder of the refining capacity (20%) is divided among approximately 275 plants owned by close to 250 companies. These companies possess about 290 cracking units.

The refining industry, like several other industries, is dominated by a score of companies. However, compared to other industries such as automotive, public utility and steel, the petroleum refining industry is divided among a relatively large number of competitive units.

Transportation by land, water and air

E. V. Rickenbacker, October 1930

In less than a generation, we must expect improvements in transportation by highways, water and air far beyond anything dreamed by the public. This could include traveling by superhighways 300 ft–400 ft wide, extending from ocean to ocean and from border to border. Commercial air travel can one-day bring hordes of passengers to faraway places and much more; the possibilities are endless.

The meaning of the gasoline distillation curve

G. Edgar, J. B. Hill and T. A. Boyd, November 1930

The purpose of this paper is to discuss the more important relationships between the distillation data of gasoline and its performance in the automobile engine. Motor fuel volatility relates to engine performance in the following ways:

- Ease of starting a cold engine
- Tendency to interrupt operation because of vapor lock
- Ease of acceleration
- Relative ease of effecting a dry mixture
- Tendency to crankcase dilution.

A study of the fundamental principles involved leads to several practical principles that may serve as a guide to oil refiners in indicating the volatility characteristics which gasoline should pos-

sess to give satisfactory engine performance under various conditions, and to the automotive engineer in indicating methods by which the fuel fee systems of automobiles may be improved.

More cracking units—Less straight-run gasoline production

C. F. Kettering, January 1931

Straight-run gasoline is fast losing its prestige. Because of its anti-knock qualities, cracked gasoline is steadily progressing in importance.

Possible use of natural gas for chemicals

H. Smith, January 1931

A significant amount of natural gas is wasted from in oil field operations/production. One possible method of utilization is in the manufacture of chemical products.

Vapor distillate stabilization and gas recovery system

A. W. Burket, March 1931

This article presents data and yields of petroleum products from various crude runs on a modern continuous vacuum distillation battery. The distillation of a given cut of oil can be carried out at an appreciably lower temperature under vacuum than under atmospheric pressure. The prime advantage of this is prevention or reduction of undesirable cracking. Therefore, heavy lubricating oils of high flash can be obtained. Another advantage is in smaller fuel bottoms yields are made possible by vacuum reduction. This means greater available yields of cracking stocks. From a mechanical standpoint, the use of vacuum by lowering distillation temperatures results in reduced deterioration of still bottoms by heavy firing.

Corrosion-proof pressure vessels

O. E. Andrus, April 1931

To help mitigate corrosion on pressure vessels, a new method has been developed which coats the entire vessel in a corrosion-proof alloy.

Progress toward a standard method for determining the anti-knock value of motor fuels

H. C. Dickinson, May 1931

Tests show that anti-knock characteristics of commercial motor fuels can be determined. These measurements are likely to be of increasing importance because of the rapid changes in motor design, and consequently in fuels, required to meet the motoring public's demand for increased power.

Waste fuels develop steam and electricity

H. R. Sharpless, July 1931

The author, Superintendent of Power, Gulf Refining Co., Port Arthur, Texas (U.S.), details the utilization of waste materials for the generation of steam and electricity. He cites some of the problems of generation and means of solution as experienced at the largest refinery in the world.

Direct cracking of crude

E. F. Nelson, August 1931

This work focuses on the cracking of light crudes when charged direct to the latest type of the Dubbs cracking unit.

Shield welded pressure vessels now safe for all hazardous uses

G. Raymond, December 1931

Of great importance are the remarkable improvements in the art of electric arc welding known collectively as "shielded welding processes." By means of these greatly advanced welding processes, joints of nearly 100% efficiency can be rapidly formed between parts of almost any thickness or shape.

Anti-knock characteristics of natural gasoline with reference to grading

R. C. Alden, February 1932

By chilling the fuel systems of knock testing engines, it is possible to determine the anti-knock ratings of gasolines of comparatively high vapor pressure. A limited number of such determinations have been made on natural gasoline and the results have been correlated with other characteristics of the gasoline. The best relationship has proven to be with Reid vapor pressure.

Skimming, cracking and reforming accomplished in one furnace

C. J. Pratt, March 1933

Simplicity is the essence of modern invention. The recently perfected petroleum refining process that incorporates simultaneous skimming, reforming, vapor phase cracking and rerunning is discussed in this article.

Octane requirements forcing cracking expansion

G. Reid, April 1932

The future of the refining industry is closely interwoven with the future progress made in the development of the cracking process. The refiner that does not possess cracking facilities will find it increasingly difficult to compete and to market their relatively low anti-knock rated motor fuels.

Integrated units controlling more refinery capacity

G. Reid, May 1932

Refining capacity continues to shift toward the integrated company. Regardless of the economics of the situation, more refining companies are engaging in the marketing of petroleum products, and more refineries are being erected by the integrated refining company.

Modern design in cracking facilities

J. J. Mack, October 1932

The trend in cracking unit design is not only toward larger capacities but includes the use of multiple furnaces, wherein two cracking coils discharging into a common secondary system lend much to flexibility of the installation. This article describes the design and operating features of this modern-style plant.

Survey of cracking plants shows continued expansion

G. R. Hopkins, October 1932

Role of sulfuric acid in the treatment of pressure distillate

A. W. Trusty, December 1932

Although various treating methods have been put into commercial operation in the last 10 yr, the sulfuric acid method of

treating petroleum naphtha continues to be the most widely used. This article provides the advantages and disadvantages of treating cracked distillate with sulfuric acid.

Design of treating system includes facilities for purging

January 1933

Preparation of ethyl alcohol from ethylene

V. Gerr, O. Pipik and E. Mezhebovskaya, February 1933

In this series of articles, information is presented dealing with the preparation of ethyl alcohol from petroleum gases both in the laboratory and in commercial scale. Ethylene was separated from the homologs by making use of the selective absorption ability of activated charcoal and separated in the form of ethyl sulfuric acid and the alcohol separated by hydrolysis. Silver and iron catalysts were used to accelerate the reaction. The constants of the synthetic alcohol coincide with those of fermentation alcohol.

Meeting some of the problems of the pressure vessel user

T. McL. Jasper, O. E. Andrus and L. J. Larson, April 1933

The history of pressure vessel development can be characterized by a series of increasing demands by the pressure vessel user, which have been met by the pressure vessel manufacturer.

The invention of the steam engine marks the starting point in the demand for pressure equipment on a quantity basis. This demand started on a relatively small scale and has expanded into the use of large quantities of steel and other materials. Many different methods of factrication have been used for producing pressure vessels. Cast iron, bolted and riveted wrought materials, forge welded cylinders, solid forgings and, most recently, autogenous welded vessels have marked the general history of the development.

Tube wall telescope for examination of interior of tubes and piping

May 1933

This article details an invention that enables plant personnel to examine the inside of tubes and piping.

Development of naphtha reforming practice for octane number improvement

E. J. LeRoi and H. W. Ferguson, June 1933

Naphtha reforming is a direct outgrowth of the ever-increasing demand for higher octane number motor fuels. Since the refiner has been dependent in the past on selection of crudes and blending fluids for anti-knock improvement, the application of naphtha cracking lends greater flexibility to refinery operations. This article details three years of experimental work on both straight-run and cracked heavy naphtha. The combined and individual effects of time, temperature and type of feedstock on yield, capacity and product quality have been carefully analyzed for varying degrees of octane number improvement.

Removal of hydrogen sulfide from high-sulfur gases

P. J. Wilson, July 1933

This article details three new processes—developed in the past year—to remove hydrogen sulfide from high-sulfur gases.

Comparison of cracking and hydrogenation as methods of producing gasoline

R. T. Haslam, R. P. Russell and W. C. Asbury, September 1933

This article focuses exclusively on the use of the hydrogenation process for gasoline production from gasoil. A critical analysis has been made of the factors which affect the economic position of the hydrogenation process, the ability of the process to compete with cracking and its use in conjunction with cracking.

Automatic temperature control in oil refining

V. R. Chadbourne and P. E. Darling, December 1933

Deviations from optimum operational conditions result in decreased yields, shortened runs and a departure from the desired product specifications. Modern refining units have variables such as flow, furnace temperatures, outlet temperatures and tower levels held within very close limits by automatic devices. Any control that will increase the yield by a fraction of one percent, even though of high initial cost, will be paid for by the increased yield in a remarkedly short time.

Design of high-pressure rectifiers

I. N. Beall, March 1934

This series of articles focuses on high-pressure rectifiers that are in use for the stabilization of natural gasoline.

Regular grade cracked gasoline and "Q" grade ethyl compared

W. Hubner and G. B. Murphy, May 1934

While the octane number method of rating gasoline quality has become generally reconfirmed and accepted by the automobile and oil industries, doubts have been growing in the minds of many technologists as to the adequacy of this method as the only criterion on which to measure the value and base the price of motor fuel. Because of these doubts, road and laboratory dynamometer tests were made in which the performance of cracked gasoline from one refinery, undoped, was compared against "Q" grade gasoline of equal volatility and equal octane rating.

The tests showed that there are desirable performance properties not shown by octane number tests, primarily miles per gallon and horsepower hours per gallon. Conclusions drawn from the road and laboratory tests showed non-ethylized regular-grade cracked gasoline is a motor fuel superior to "Q" grade ethyl of equal volatility and equal octane number rating.

Fouling of heat exchangers

W. L. Nelson, July 1934

These articles focus on the fouling of heat exchangers and the fouling factors which are obtained in classes of heat transfer equipment. Part 1 provides an introductory to heat transfer and fouling, with Part 2 focusing on fouling factors that can be expected in plant service, the deposition of coke in pipe still tubes, the temperatures that are attained in pipe still tubes and the fouling conditions caused by hard water and wax.

Prospects of a petroleum chemical industry

C. Ellis, September 1934

During recent years, there has been a steady trend toward the establishment of a chemical industry dependent upon petroleum as a source of raw material.

The efficiency of petroleum fractioning column

V. W. Garton and R. L. Huntington, January 1935

Separation of crude oil into its several cuts by means of fractional distillation is one of the most important processes used in the petroleum refining industry. This process depends on four factors:

1. The number of bubble plates
2. The molar ratio of reflux, or overflow to the rising stream of vapor
3. The approach toward equilibrium conditions between liquid and vapor of each plate
4. The amount of entrainment or mechanical carrying of liquid droplets or mist from one plate to another.

This article provides a simple method to determine the fractionation capacity of any laboratory packed column, a knowledge of which is essential in the analysis of fractions from commercial towers and in the calculation of their efficiencies.

The trend in design, construction and operation of gasoline plants

J. Campbell, March 1935

The author states, "The modern natural gasoline plant is quite different from the plant of yesterday, but the plant of tomorrow will be better arranged, even more completely automatic—if such an expression may be used—and less wasteful, as well as more efficient and safer. However, we will have the personal element to contend with as long as human beings must operate and watch over operations. Each of us who has any responsibility for the construction and operation of these plants should adopt as their personal motto and pass on to all those under their supervision this slogan, 'eternal vigilance is the price of safety'."

Inspection of oil refinery equipment

F. Newcomb, April 1935

Inspections should not be conducted by those directly responsible for the production or maintenance of equipment. Thorough inspections should be conducted by independent forces, with the freedom to state conditions exactly as found and to criticize the condition of the equipment without fear of jeopardizing their positions. This structure can lead to optimal operation of refining equipment.

Fuel specifications for high-speed Diesel engines

G. C. Wilson, June 1935

The development of the high-speed Diesel engine depends as much upon proper fuel specifications as upon engine design. Therefore, refiners and engine builders will find it advantageous to cooperate in solving fuel problems.

The various items included in specifications are discussed here. The importance of ignition quality and its effect on engine operation are emphasized.

Cost of steam and heat in the refinery

W. L. Nelson, April 1936

Steam and heat are the most important items in refining operations and the proper utilization, conservation and distribution of steam and heat influences the profit and loss of a facility. This series of articles focuses on the economics of steam and heat usage within a refinery.

Instrumentation in oil refining

A. C. Proctor and G. Egloff, June 1936

One of the important developments in the processing of oil has been in automatic control. This development has brought about marked economic savings due to producing better and higher yields of the more valuable products from crude oil.

Natural gas as a chemical raw material

I. N. Beall, July 1936

Natural gas consists of relatively few components, and these are easily separable in substantially pure form by methods already established. In most chemical processes, the purity of the basic raw materials is of considerable importance. In this article, the author examines processes that could utilize natural gas for chemicals production.

Salt removal from crude oil— Chemical and physical methods

E. R. Jones, May 1937

This article discusses the observations made and the results obtained while trying to find an immediate practical solution for removing salts from crude oil.

Cathodic protection: Its application in refineries

D. S. Sneigr, July 1938

Cathodic protection of metals has been found worthwhile in protecting pump shafts. Refinery and other tank farms use it for protection of large tank bottoms, and its use has extended to all types of pipelines for carrying oil, natural gas, refinery gas, water and many others. This article reviews the development of cathodic protection and describes the various types of equipment and systems available and in wide usage.

Lower paraffins over activated alumina catalysts

J. Burgin, H. Groll and R. M. Roberts, October 1938

Increasing demands for lower olefins as base materials for synthetic gasoline and chemical products have led to the investigation of catalytic dehydrogenation processes for their production from lower paraffins. This article provides results of experiments with activated alumina, alone and combined with chromium oxide, which has proved to be a selective catalyst for dehydrogenation.

Preparation of blending stock for 100-octane gasoline

L. J. Coulthurst, March 1939

Aviation gasoline having a 100 octane anti-knock rating has become established as an economic reality within the last 2 yr and is no longer regarded as a laboratory curiosity. There is no doubt that commercial planes will be using 100-octane motor gasoline exclusively in the very near future. HP

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LEE NICHOLS, Editor-in-Chief/Associate Publisher

Polymer science, catalytic cracking and chemical advances

HERMAN FRANCIS MARK

The Austrian-born chemist, Hermann Francis Mark, is well-known for his contributions to the development of polymer science, which he devoted more than 60 yr of his life to. While working with IG Farben in Germany, Mark worked on experiments on the commercialization of polymers such as polystyrene, polyvinyl chloride and the first synthetic rubbers.[1]

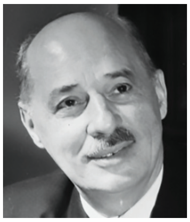
After escaping Nazi Germany, Mark found his way to the U.S. and started classes on polymers at the Polytechnic Institute of Brooklyn, later founding the Polymer Research Institute, which was the first facility devoted to polymer research. For his lifetime of work, he received the U.S. National Medal of Science in 1979.

OTTO RÖHM

Otto Röhm was a German chemist and pharmacist that founded Röhm and Hass AG. His experiments with methyl methacrylate (MMA) led to the development of Plexiglas.

After successfully developing and marketing Oropon, a more hygienic and efficient way of staining leather, Röhm focused his sights on plastics research. While working with Walter Bauer, researchers conducted an experiment polymerizing MMA between two layers of glass in a water quench. The result was a clear plastic sheet that was lighter than glass but much less prone to shatter. The material, called Plexiglas, would first be used as a substitute for glass in military aircraft, eventually being used in many industrial and commercial applications.

EUGENE HOUDRY



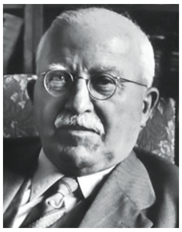
With the aide of E. A. Prudhomme, French engineer Eugene Houdry is known as a pioneer in catalytic cracking. After serving in WWI in the French artillery division and later in the tank corps, Houdry worked in his father's steel business, as well as raced cars. His passion led him on a pathway to improving engine performance.

Prior to Houdry's discovery, thermal cracking was the primary refining process to produce gasoline. However, many researchers and analysts feared that thermal cracking was insufficient to satisfy increasing global demand for gasoline. Houdry and Prudhomme's research led to the development of the fixed-bed catalytic cracking unit. Operations of the 15,000-bpd unit began at Sun Oil's Marcus Hook refinery in Pennsylvania (U.S.) in 1936. Approximately 50% of the 15,000-bpd unit produced high-octane gasoline, which was double the production of conventional thermal processes.[2] The novel process produced high-octane gasoline—the Houdry unit could produce 100-octane aviation gasoline, which provided U.S. military aircraft a significant advantage over Germany.

OTTO BAYER

While working at IG Farben in the late 1930s, German chemist Otto Bayer conducted extensive polymer research that led to the discovery of polyurethane. One such experiment created a new polymer by reacting 1,8 octane diisocyanate with 1,4 butanediol. This new polymer, polyurethane, was first used as coatings and adhesives. It was a suitable replacement for rubber during World War 2 (WW2). Post-WW2, the product was used extensively in many applications, and is still widely used today. This includes in insulation, building materials, adhesives, coatings and clothing, among others.

HERMANN STAUDINGER



The Nobel Prize-winning German chemist is best known for his research on macromolecules, which he characterized as polymers. Staudinger also discovered ketenes, which would later be used to produce antibiotics.[3]

Staudinger hypothesized that polymers were linked end-to-end. His work with high-molecular weight compounds provided the foundation for polymer chemistry. He authored hundreds of scientific papers and several books on topics such as macromolecular chemistry and biology. His research on macromolecular chemistry earned him a Nobel Prize in 1953.

WALLACE CAROTHERS

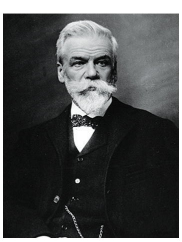
The American chemist started work at DuPont in the late 1920s. His primary focus was on polymer research. Under his tenure, DuPont would produce several long-lasting discoveries that would revolutionize the chemical industry.

In 1930, Carothers and his staff conducted experiments and research on an acetylene polymer. The goal of the research was to create synthetic rubber. After several tests, the group produced a substance that resembled rubber, which later took the name Neoprene.

Carothers' group was also credited with producing the first synthetic silk. This synthetic polymer would later be called polyester, which is still in use today.

By the mid-1930s, Carothers produced fibers comprised of amine, hexamethylene diamine and adipic acid. These new strong, elastic fibers were called polymer 6,6 (or nylon 66). Nylon first became a household product as women's hosiery, later being used in the U.S. war effort to produce parachutes and tents. Over the next several decades, nylon would be used extensively as a combined fabric in fashion and apparel, as well as in several industrial applications—the global nylon industry market size is forecast to reach more than $46 B by the late 2020s.[3,4]

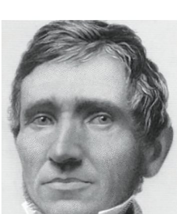
ERNEST SOLVAY



The Belgian chemist and industrialist is known for developing the ammonia-soda process to manufacture soda ash on a commercial scale. The process was invented by Ernest and his brother Alfred in the early 1860s. In 1863, the brothers founded Solvay and Cie, opening their first soda ash plant in Couillet, Belgian shortly thereafter.[5]

Soda ash was widely used in several industrial applications. The wide use of the material enabled the Solvay brothers to expand operations into other countries, such as Austria, Germany, Russia, the UK and the U.S. By 1900, 95% of soda ash consumption around the world was produced by the Solvay process. Many of these plants are still in use today.

CHARLES GOODYEAR



The American self-taught chemist is known for developing volcanized rubber, which revolutionized the industry. Goodyear's dive into better rubber materials began while visiting the Roxbury India Rubber Co. in New York (U.S.). After examining life vests, he believed he could improve the valves on the vests. However, the store manager made the comment to Goodyear that he would be better off inventing a better rubber. [6]

Over the next several years, Goodyear worked tirelessly on developing better rubber, even going nearly bankrupt in the process. However, while working at the Eagle India Rubber Co., Goodyear accidentally discovered the vulcanization of rubber by combining rubber and sulfur over a hot stove.[6] Once heated, the rubber hardened. In 1844, he finally perfected the process and was given a patent for his invention—the process was called vulcanization after Vulcan, the Roman god of fire.[6] His work led to the development of a vulcanized rubber producing hub in the northeast U.S., leading to the Goodyear company being named in his honor in the late 1890s. HP

ACKNOWLEDGEMENTS

Hydrocarbon Processing would like to thank several institutions/companies for the use of archived images of industry pioneers. These include Michigan State University's College of Natural Science Department of Chemistry, Röhm and Hass AG, Science History Institute, Bayer, Solvay and Goodyear.

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P. DIXIT, Consultant, Baroda, India

Construction pitfalls in SRU-fabricated equipment

In the hydrocarbon processing industry, sulfur recovery is a well-known process. However, this process uses a variety of equipment. Due to the number of construction activities being carried out onsite, this work requires special attention and skill. This article covers lessons learned by the author while working on sulfur recovery units (SRUs). Many of the pitfalls discussed could have been avoided if supervision and site inspections were more vigilant. The equipment under consideration includes the reaction furnace, the waste heat boiler (WHB), sulfur condensers, Claus converters, sulfur locks, incinerators and stack. Sharing this information may be useful to engineers working on similar equipment.

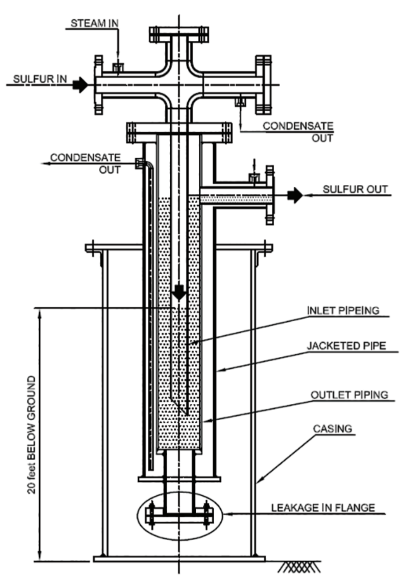

FIG. 1. Leakage in the sulfur lock.

Choked pipe and leaking sulfur seal locks. The Claus unit converts hydrogen sulfide (H_2S) into liquid sulfur. The liquid sulfur from the condensers passes through the sulfur locks (**FIG. 1**), which provides a U-tube seal to stop the reverse flow of H_2S. The equipment is tall (7 m–8 m) and is placed in a casing underground, making installation and removal a rigorous job.

In one case, the seal locks choked, causing backflow into the condenser. This led to the seal lock casing filling with sulfur, which overflowed, resulting in a shutdown. The seal lock was then removed and examined. It was discovered that the bottom was full of debris. The reason may have been that the upstream equipment and piping were not thoroughly cleaned prior to startup.

In addition, it was noticed that the bottom flange of the seal lock had a leak. The bottom flange was composed of a plate flange with a raised face. The flange face was likely damaged due to overtightening during a hydrotest. The pair of flanges were replaced by a weld neck, a ring joint flange (which is much stronger in design) and a ring gasket.

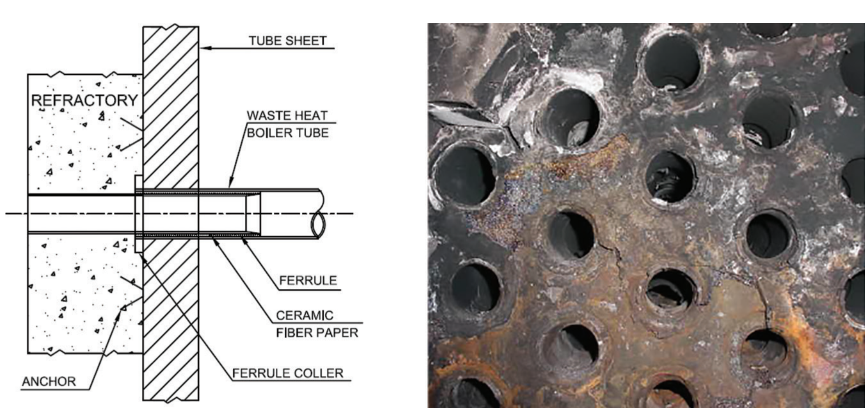

FIG. 2. Cracks in ferrules and the refractory.

Cracked refractory in a waste boiler tubesheet and broken ferrules. WHBs and sulfur condensers recover heat from flue gas and generate steam. The steam is generated on the shell side. The tubesheet that faces the reaction furnace is subject to a considerable amount of radiation and convection heat. This is applied with refractory, which keeps the tubesheet temperature low and protects the tube-to-tubesheet joint. Ferrules (ceramic inserts) are used at the end of the tube to transfer heat flux away from the tubesheet joint.

After curing of the refractory, cracks were noticed in the refractory and part of the ceramic ferrules were found broken (**FIG. 2**). Investigation revealed that there were a few unbroken ferrules in good condition. It was decided to break these manually to determine the reason of their survival.

As per design/drawings, each ferrule was supposed to have ceramic paper

wrapped around its surface before being inserted into the tube. This ceramic fiber paper keeps the ferrule flexible to move within the tube and does not allow the refractory to stick to the tube.

The ceramic fiber paper was found in unbroken ferrules but was missing in cracked pieces. After checking with the field supervisor, it was discovered that the ceramic fiber paper stock fell short, and workers went ahead without it. Therefore, the ferrules bonded with the inner wall of the tube during application of the refractory and during curing. Since the ferrules were rigidly bonded, they broke during curing.

Refractory anchors in furnaces. Refractory anchors play an important role in holding the castable refractory during hot and cold conditions.

Various refractory anchor designs and shapes are used for different applications. These anchors are welded to the furnace body at a specified pitch. The welding pattern and pitch of these anchors differs, depending on the refractory thickness, weight and orientation in the furnace. A strong anchoring system is key to maintaining monolithic refractory lining integrity—even when it is cracked—to prevent a total structural collapse.

The tips of the anchors experience the highest temperatures. To allow expansion of the anchor tips, they are covered with plastic caps or dipping wax. The tip burns out, allowing the anchor to expand without causing cracks in the refractory.

The following are common mistakes/omissions regarding anchors and related field work:

- **The mix-up of anchors:** Mix-up of different metallurgy or type.
- **Anchor pitch or orientations not followed:** Anchor pitch varies, based on the anchor's location in the side furnace. Sometimes, a worker misses this aspect and starts welding on the uniform pitch and orientation.
- **Anchor weld joint hardness:** Welding the anchor, using high heat input, makes the joint brittle. Since these are welded sequentially, the work is easy. However, if any anchor breaks in between, repair is quite difficult. A typical mock-up should be made and tested by knocking the anchor, adjusting the welding current and making a procedure for each type. The joint should be ductile and not break.
- **Missing anchor expansion caps:** Any missing caps should be put in place prior to application of refractory.

Refractory drying out. The WHB tubesheet ferrule design is important to protect WHB tubes. If a crack develops, hot process gas may reach the tubesheet and/or tubes, resulting in high-temperature sulfide corrosion and eventually tube failure.

Refractory curing heating rates are normally provided by refractory vendors. Speeding up the commissioning process results in refractory failures, leading to hot spots on the metal shell and breakages in refractory.

Refractory linings must be dried very slowly prior to the introduction of acid gas to ensure that moisture is completely removed. If not properly dried, embedded moisture may form steam, resulting in a crack in the refractory (**FIG. 3**).

Catalyst slippage from the converter. Converters are horizontal vessels that support catalysts on bar-type gratings (**FIG. 4A**). Normally, the top of the grating contains stainless-steel mesh held under the weight of ceramic balls, with the catalyst on the top.

On a project, catalyst leakage was found in large quantities in the downstream sulfur condensers. On investigation, it was noticed that the wire mesh above the grid did not cover the grating; it had slipped out from its place, creating an opening near the side wall.

To overcome this, the design was modified. An overlap of 4 in.–6 in. of mesh on the vessel wall was used and held in position by additional stud anchors and a clamping plate (**FIG. 4B**).

Sulfur condensers (mist eliminators and inclined nozzles). Sulfur condensers are conventional shell-and-tube heat

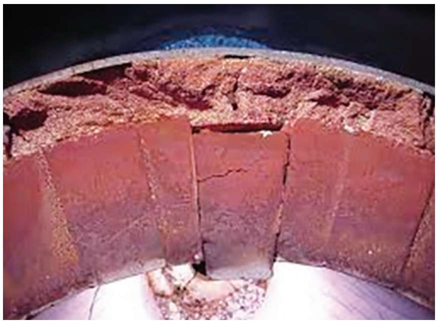

FIG. 3. Refractor damage caused by dry-out.

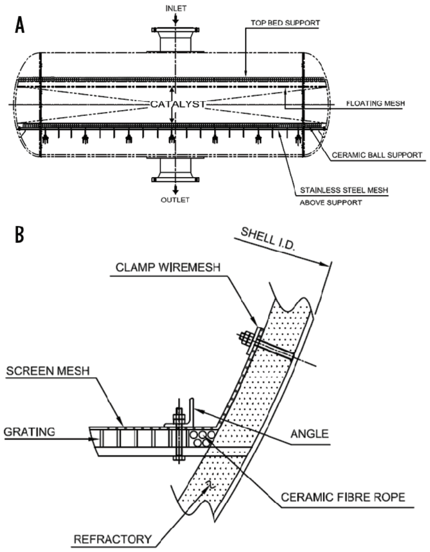

FIG. 4. View of converter mesh (A) and clamping plate (B).

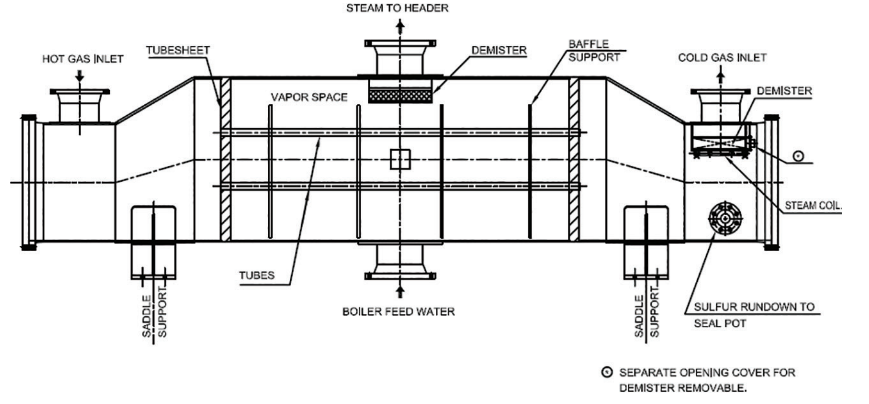

FIG. 5. Modified sulfur condenser demister box.

exchangers (usually kettle type) with some special design requirements. Medium-pressure or low-pressure steam is generated on the shell side, while liquid sulfur flows on the tube side as sulfur vapors are cooled below the dewpoint.

Both the steam side and channel side of the condenser are provided with a mist eliminator (demister). The shell side is clean service (water and steam); therefore, the shell-side demister does not need frequent cleaning.

The channel-side wire mesh type demister is kept hot by a steam coil surrounding the mesh pad to avoid sulfur condensation. In one design, it was found that the mesh pad was blocked from all sides. Therefore, without modification, it was impossible to remove the demister. **FIG. 5** provides details of the modified demister box.

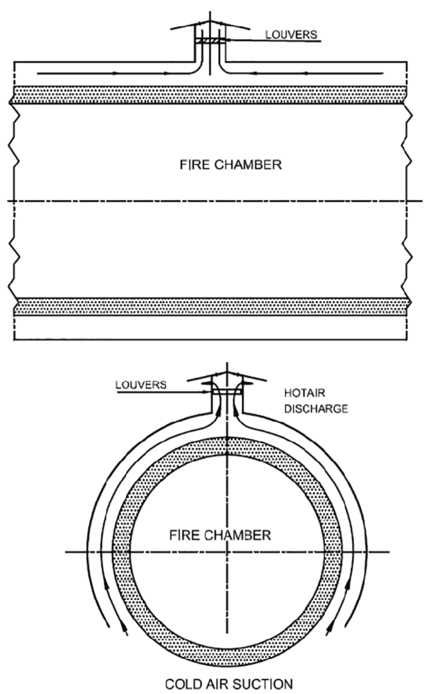

FIG. 6. Shroud and louvers for a furnace.

Shrouds and louvers. The main reaction furnace is a vital piece of equipment in a sulfur plant. The high temperatures and operating conditions in the burner and furnace require protection of the carbon-steel body metal via a thermal shroud and, internally, via an external and multilayered refractory lining.

The system is designed to maintain the inner-side metal temperature between 150°C and 300°C to avoid both excessive acid and sulfide corrosion. The shroud also handles external temperature variations during winter and rains, as it maintains an air gap around the shell. The gap is designed to control air flow around the equipment to maintain the temperature of the metal. For this purpose, variable openings are provided at the top of the shroud with louvers and hats (**FIG. 6**). The louver openings can be manually or automatically adjusted to control convective air flow. The opening size and operability of louvers must be checked and inspected by a field engineer before plant commissioning. Sometimes, louver openings differ from drawings and are inoperable.

Overheating of the thermal oxidizer. All tail gases are burned in an incinerator at the end of a process.

A case was encountered where, during summer days, the operator complained of very high temperatures outside and around the incinerator. It was not possible to get closer to the equipment for operation and maintenance, and there was also the chance of damaging delicate instruments and cables near the equipment.

Rechecking the design and thermal calculations, it was noticed that:

- The original design was for a unit in a colder country.
- The insulating castable refractory used was below specification, resulting in an increase in the vessel's body temperature.

The challenges were resolved by replacing the old refractory with one of superior quality.

Wrong location of a guide support for a steel stack. The final effluent gas from the process is incinerated in an incinerator. In a specific case, it was a small-diameter, 12-in. stainless-steel stack, with a height of around 25 m and a guide support at two levels. A supporting structure was used. One support was in the middle and another was near the top. The guides were a fin-type support (**FIG. 7**) that passed through the structure's frame opening.

In this incident, everything went well during startup, commissioning and operation. However, when a shutdown was taken, it was noticed that the top guide structure had been slightly pulled down by the stack.

After an investigation, it was discovered that the guide fin lengths were not placed at the right elevation to handle the stainless-steel pipe expansion. During operations, the stack expanded upward, and the guide supports went above the structure. It was resting on a beam structure during operation and, due to the wind load pressure on one side, got stuck. During shutdown, the stack cooled and shrunk, causing the guide fins to pull down the structure. HP

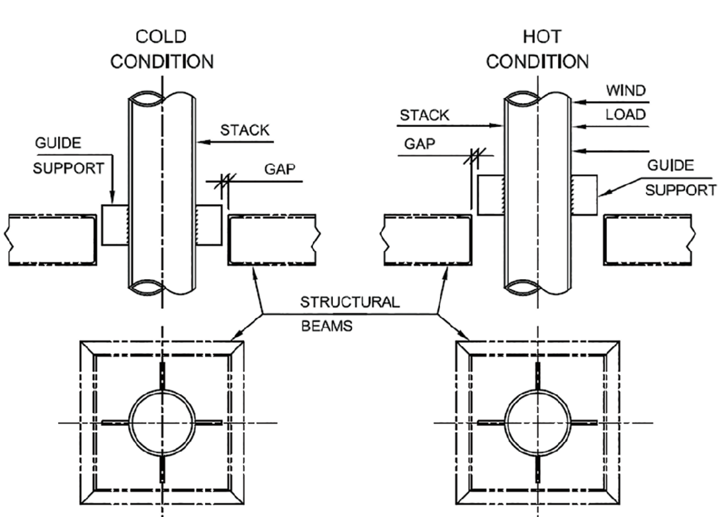

FIG. 7. Guide support for a vent stack.

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Calculate temperature at the bottom of pipe supports and heat loss rate

Pipes transporting hot media are insulated to reduce heat loss and maintain the required process conditions. Shoe or trunnion supports are used to support such pipes at regular intervals. Since the support is directly connected to pipe, there will be conductive heat transfer through the support and pipe. This creates an uninsulated heat bridge between the pipe and supporting structure via the support shoe or trunnion, as shown in **FIG. 1.**

For high-temperature lines (≥ 400°C), a substantial amount of heat transfer can occur through the pipe support. As a result, significant temperatures can be observed on the structure surface (bottom of shoe or trunnion). Engineering consultants and clients should limit the pipe support bottom temperature to < 350°C resting on steel and < 150°C resting on concrete to avoid overstressing/degradation of the supporting beams.

Therefore, for high-temperature lines, it is vital to establish a shoe height that complies with temperature limits at the pipe support. If the shoe height cannot be increased adequately, a decision can be made to put a heat isolation block at the bottom of the pipe support.

Computational fluid dynamics (CFD) or finite element analysis (FEA)-based simulation software can be used to calculate temperature at the bottom of the pipe support. However, CFD/FEA are labor intensive, adding to the project cost and time delays considering the substantial number of pipe supports in a plant.

This article provides a method to analyze heat transfer through pipe supports and calculate temperatures at the pipe support bottom. Based on theoretical basis explained in subsequent sections, a mathematical model was developed to determine the temperature at the bottom of a pipe support, as well as heat loss rate. Calculation results are supported by results performed using CFD software.[1,2,3,4]

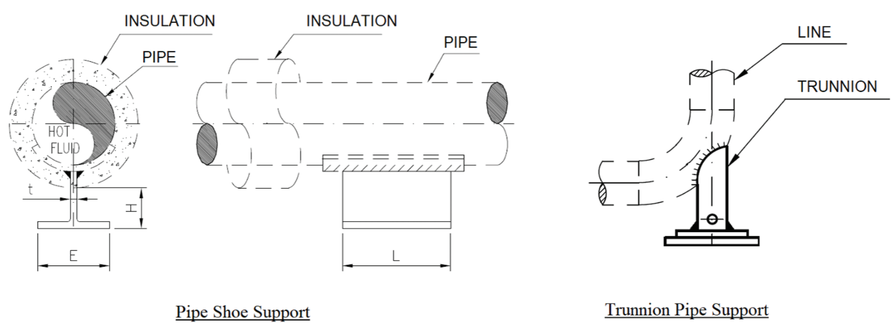

FIG. 1. Pipe support geometry.

Theoretical basis. Heat transfer occurs from the hot pipe to the structure via the pipe support through various modes of heat transfer, including (**FIG. 2**):

a. Heat conduction between the inner and outer layers of the pipe
b. Heat gain from the source (i.e., pipe) or conduction through the pipe support
c. Atmospheric cooling or convective heat loss through the pipe support
d. Radiative heat loss through the pipe.

From the above, mechanisms (b) and (c) are more dominant in piping scenarios.

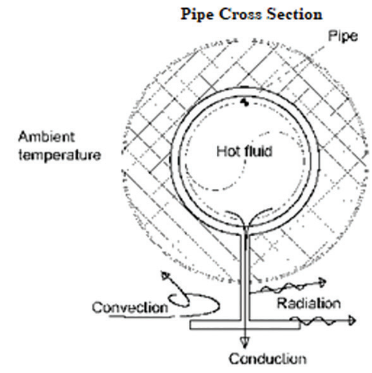

FIG. 2. Modes of heat transfer in a pipe shoe support.

For the discussion here, a list of unit abbreviations is included:

t = Thickness of shoe/thickness of trunnion support, mm
H = Height of pipe support (shoe or trunnion) outside pipe insulation, mm
E = Width of bottom plate for shoe support, mm
L = Length of shoe support, mm
D = Outside diameter of trunnion, mm
P = Perimeter of pipe support (shoe or trunnion), mm
A_c = Cross-sectional area of pipe support that is exposed to atmosphere/outside of insulation, m^2
T_b = Pipe temperature, K
T_H = Temperature at the bottom of the pipe support, K
T_∞ = Ambient temperature, K
h = Convective heat transfer coefficient for air, W/m^2-K

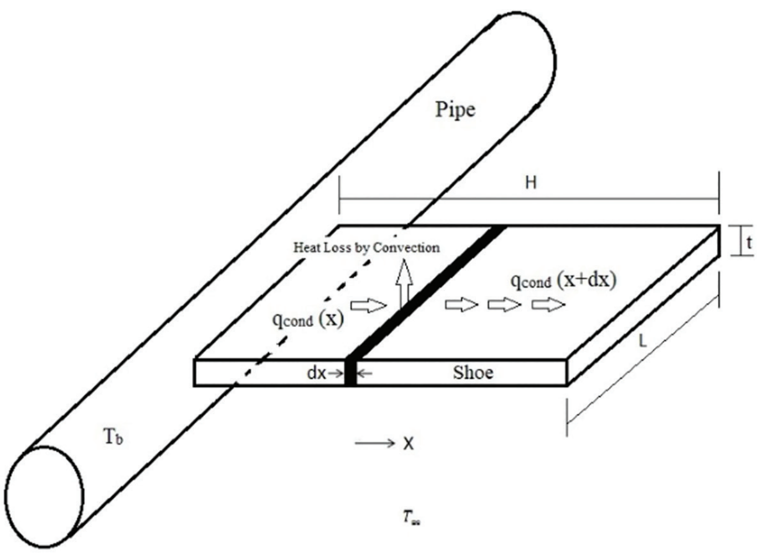

FIG. 3. Energy balance over a small element of the shoe support.

k = Conductive heat transfer coefficient for pipe support material, W/m²-K
x = Length variable
$cos\,h$ = Hyperbolic cosine function
Re = Reynolds Number
Pr = Prandtl Number
Nu = Nusselt Number
ρ = Air density, kg/m³
v = Air velocity, m/sec
μ = Dynamic viscosity of air, N-sec/m²
C_p = Specific heat of air constant pressure, J/kg-K
k_{air} = Thermal conductivity of air, W/m-K
$\dot{Q}$ = Heat removal rate from the pipe through conduction, W

Mathematical model. To analyze the heat transfer, the pipe support is considered as a projection/fin that always remains in contact with the pipe. When the pipe is supported with a shoe, the cross-section of the projection will be rectangular. Alternatively, the cross-section will be circular in the case of a trunnion support. The pipe and support junctions are considered as the base, and the temperature of the base is already known. The heat from the pipe is transferred through the support via conduction. As the support is exposed to atmosphere, heat loss through convection takes place simultaneously.

As shown in **FIG. 3.** the energy balance/energy conservation across a small element of the pipe support (shoe type) can be written as Eq. 1:

$$\begin{pmatrix}\textit{Heat received at distance}\\ (x+dx)\ \textit{through conduction}\end{pmatrix} = \begin{pmatrix}\textit{Heat received at distance } x\\ \textit{through conduction}\end{pmatrix} + \begin{pmatrix}\textit{Heat loss to the atmosphere}\\ \textit{via convection through length } dx\end{pmatrix} \quad (1)$$

Or, $q_{cond}(x+dx) = q_{cond}(x) + q_{conv}$

Using Fourier's Law of heat conduction (Eq. 2):

$$q_{cond}(x) = -kA_c\left(\frac{dT}{dx}\right) \quad (2)$$

and Newton's Law of cooling (Eq. 3):

$$q_{conv}(dx) = Ph(T - T_\infty)dx \quad (3)$$

Plugging in the values (Eq 4):

$$-kA_c(dT/dx)_{x+dx} = -kA_c\,(dT/dx)_x + Ph(T - T_\infty)dx \quad (4)$$

This expression can be rearranged into a final form for a uniform cross-section (Eq. 5):

$$\frac{d^2T}{dx^2} = \frac{hP(T-T_\infty)}{kA_c} \quad (5)$$

This differential equation can be solved for the following two boundary conditions (Eqs. 6 and 7):

$$T\big|_{x=0} = T_{pipe} \quad (6)$$

The other end of the shoe plate is insulated (as the cross-section is small at the other end, the heat loss through it is negligible):

$$\frac{dT}{dx}\Big|_{x=H} = 0 \quad \text{or} \quad q_{conv}\big|_{x=H} = 0 \quad (7)$$

Simplified, this results in a solution (Eq. 8):

$$\frac{T-T_\infty}{T_b-T_\infty} = \frac{\cosh m(H-x)}{\cosh mH} \quad (8)$$

where (Eq. 9):

$$m = \sqrt{\left(\frac{hP}{kA_c}\right)} \quad (9)$$

Since the objective is to determine temperature at the bottom of the pipe support, we can assume $x = H$, providing Eq. 10:

$$\frac{T_H-T_\infty}{T_b-T_\infty} = \frac{1}{\cosh mH} \quad (10)$$

The amount of heat removed from the pipe represents the heat loss through the pipe support. This can be calculated using Fourier's law and the final expression is given in Eq. 11:

$$\dot{Q} = \sqrt{hPkA_c} \times (T_b - T_\infty) \times \tanh mH \quad (11)$$

Conclusively, Eq. 10 can be used to find the temperature at the bottom of the pipe support. Heat loss rate through the pipe support can be calculated using Eq. 11. The required inputs for the calculation are T_b (inner temperature for a conservative case) and h. Values of h depend on the pipe support geometry and air velocity. In this work, different strategies have been used to compute h for the shoe support and trunnion support. For a shoe support, the value of h is calculated using the following relation between convective heat transfer coefficient and air velocity (Eq. 12):

$$h = 1.16(10.45 - v + 10\sqrt{v}) \quad (12)$$

where v is the air velocity. This relation is valid in an air velocity range of 2 m/sec–20 m/sec.

Alternatively, for a trunnion support, the value of h is calculated using the Churchill relation given by Eq. 13, which is valid for forced convection due to cross-flow over a cylinder:

$$Nu = 0.3 + \frac{0.62\,Re^{1/2}\,Pr^{1/3}}{\left[1+(0.4/Pr)^{2/3}\right]^{1/4}} \quad (13)$$

$$\left[1 + \left(\frac{Re}{282000}\right)^{5/8}\right]^{4/5} \quad \textit{if } Re \times Pr \geq 0.2$$

where, $Re = \dfrac{\rho v D}{\mu}$; $Pr = \dfrac{\mu C_p}{k_{air}}$; $Nu = \dfrac{hD}{k_{air}}$

k = Conductive heat transfer coefficient for pipe support material

P = Perimeter of pipe support cross-section: $P = 2\,(t + L)$ for a shoe support, and $P = \pi \times D$ for a trunnion support

A_c = Cross-sectional area of pipe support: $A_c = t \times L$ for a shoe support, and $A_c = 0.25\pi \times [(D^2 - (D - 2t)^2]$ for a trunnion support

H = Height of pipe support outside insulation that is exposed to the atmosphere.

SAMPLE PROBLEMS

Implementation steps include:

1. Find the thermal conductivity (k) for pipe support material at pipe temperature
2. Find the convective heat transfer coefficient for air (h) as per support geometry and average air velocity in the area
3. Evaluate P and A_c as per pipe support geometry and cross-section
4. Calculate T_H using Eq. 10
5. Calculate heat loss rate through the pipe support ($\dot{Q}$) using Eq. 11
6. Consider an error margin of 5%–10% before final judgment.

Problem 1. In this section, the heat transfer through a pipe shoe support (**FIG. 4**) has been analyzed using the proposed method. The same problem was simulated using CFD software—the results obtained from both methods have been compared.

The problem description includes:

- Shoe material: Carbon steel
- $T_\infty = 22°C$

Solution implementation steps:

1. For shoe material, carbon steel @ 788 K, $k = 39.78$ W/m-K
2. For an air velocity of 2 m/sec in a forced convection case (Eq. 14):

$$h = 1.16(10.45 - v + 10\sqrt{v}) \quad (v = 2 \text{ m/sec}) \text{ or } h = 26.2069 \text{ W/m}^2\text{-K} \tag{14}$$

where the shoe height H = 150 mm or 0.15 m.

3. Given a shoe thickness t =10 mm and shoe length L = 200 mm (Eqs. 15 and 16):

$$P = 2\,(t + L) = 420 \text{ mm} = 0.42 \text{ m} \tag{15}$$

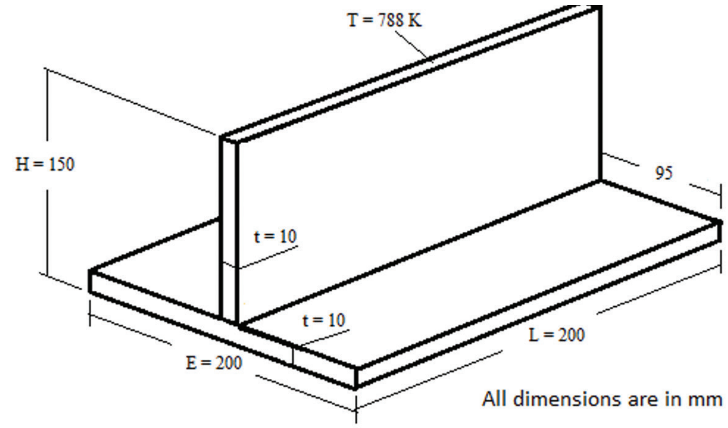

FIG. 4. Shoe support geometry for Problem 1.

$$A_c = t \times L = 10 \times 200 \text{ mm}^2 = 0.002 \text{ m}^2 \tag{16}$$

4. Using Eq. 10, we have Eq. 17:

$$\frac{T_H - T_\infty}{T_b - T_\infty} = \frac{1}{\cosh mH} = \frac{1}{\cosh\sqrt{\left(\frac{hP}{kA_c}\right)}H} \text{ or }$$
$$\frac{T_H - 295}{788 - 295} = \frac{1}{\cosh mH} = \frac{1}{\cosh\sqrt{\frac{26.2069 \times 0.42}{39.78 \times 0.002}} \times 0.15} \tag{17}$$

On solving, we get Eq. 18:

$$T_H = 459.1 \text{ K or } 186.1°C \tag{18}$$

On solving the same problem using CFD software, the average temperature at the bottom of shoe support is (Eq. 19):

$$T_H\big|_{CFD} \approx 460\ K \text{ or } 187°C \tag{19}$$

Similarly, the temperature at the bottom of the shoe support has been calculated at various air velocities, and the results have been compared with CFD simulation results.

5. For the conductive heat loss rate through the pipe shoe, using Eq. 11, we calculate (Eq. 20):

$$\dot{Q} = \sqrt{\frac{26.2069 \times 0.42 \times}{39.78 \times 0.002.}}(788 - 295). \tanh\sqrt{\frac{26.2069 \times 0.42}{39.78 \times 0.002}} \times 0.15 \tag{20}$$

or $\dot{Q} = 435.04$ W

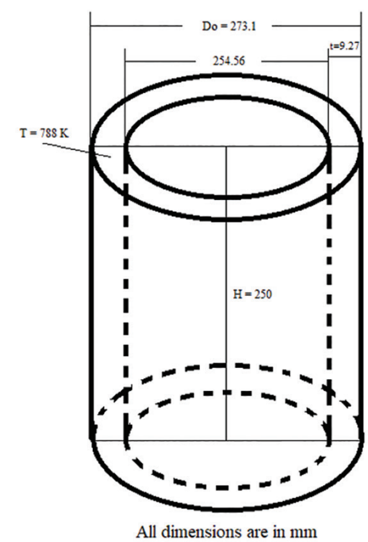

FIG. 5. Tunnion support geometry for Problem 2.

Problem 2. In this section, the heat transfer through a 10-in. STD trunnion pipe support (**FIG. 5**) has been analyzed using the proposed method. The same problem has been simulated using CFD software and the results obtained from both methods have been compared.

The problem description includes:

- Trunnion material: Carbon steel
- T_∞ =22°C
- Air density (ρ) = 1.225 kg/m^3
- D = 0.2731 m
- Dynamic viscosity of air (μ) = 1.7894×10^{-5} N-sec/m^2
- Specific heat of air constant pressure (C_p) = 1006.43 J/kg-K
- Thermal conductivity of air (k_{air}) = 0.0242 W/m-K

Following the solution implementation steps:

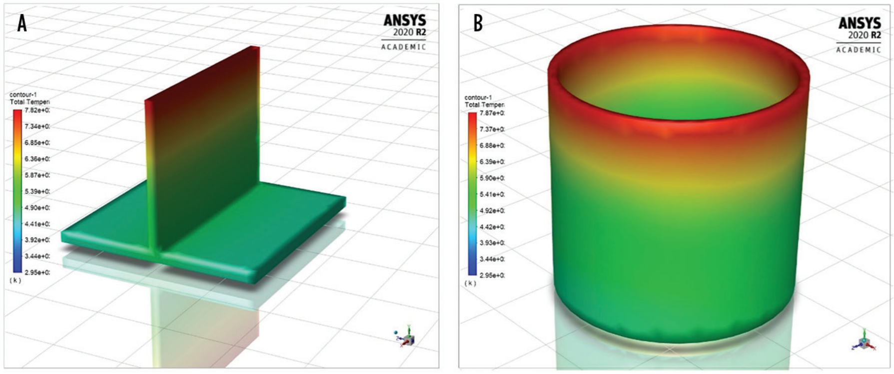

FIG. 6. Temperature contours for (a) shoe support and (b) trunnion support.

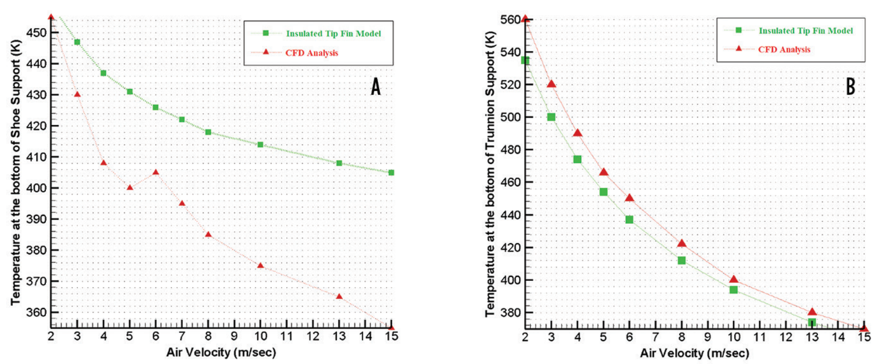

FIG. 7. Comparison of results (average temperature at the bottom of the pipe support) with CFD simulation results for (a) shoe support and (b) trunnion support.

1. For trunnion material, carbon steel @ 788 K, $k = 39.78$ W/m-K
2. For an air velocity of 2 m/sec in a forced convection case (Eq. 21):

$$Re = \frac{\rho v D}{\mu} = \frac{1.225 \times 2 \times 0.2731}{1.7894 \times 10 - 5} = 37392.14$$

$$Pr = \frac{\mu C_p}{k_{air}} = \frac{1.7894 \times 10 - 5}{0.0242} = 0.74418$$

$$Re \times Pr = 27826.33 \left(\begin{array}{l} >0.2; \\ Churchill\ equation \\ is\ appliable \end{array} \right)$$

$$Nu = 0.3 + \frac{0.62 Re^{1/2} Pr^{1/3}}{\left[1+(0.4/Pr)^{2/3}\right]^{1/4}} \left[1+\left(\frac{Re}{282000}\right)^{5/8}\right]^{4/5} = 117.1031$$

$$Using\ Nu = \frac{hD}{k_{air}}; \quad (21)$$

$$gives\ h = \frac{Nu.k_{air}}{D} = 10.3768$$

where the trunnion height H = 250 mm or 0.25 m.

3. Given trunnion thickness t = 9.27 mm (Eqs. 22 and 23):

$$P = \pi \times D = 857.534 \text{ mm} = 0.857534 \text{ m} \quad (22)$$

$$A_c = 0.25\pi \times \{(D^2 - (D - 2t)^2\} = 0.0076795 \text{ mm}^2 \quad (23)$$

4. Using Eq. 10, we calculate Eq. 24:

$$\frac{T_H - T_\infty}{T_b - T_\infty} = \frac{1}{\cosh mH} = \frac{1}{\cosh \sqrt{\left(\frac{hP}{kA_c}\right)} H} \quad or \quad (24)$$

$$\frac{T_H - 295}{788 - 295} = \frac{1}{\cosh mH} = \frac{1}{\cosh \sqrt{\left(\frac{10.3768 \times 0.857534}{39.78 \times 0.0076795}\right)} \times 0.25}$$

On solving, we get Eq. 25:

$$T_H = 534.7 \text{ K or } 261.7°\text{C} \quad (25)$$

On solving the same problem using CFD software, the average temperature at the bottom of the trunnion support is (Eq. 26):

$$T_H|_{CFD} \approx 560\ K \text{ or } 287°\text{C} \quad (26)$$

Similarly, the temperature at the bottom of the shoe support has been calculated at various air velocities and the results have been compared with CFD simulation results.

5. For the conductive heat loss rate through the pipe shoe, Eq. 27 is used:

$$\dot{Q} = \sqrt{10.3768 \times 0.857534 \times 39.78 \times 0.0076795.}$$

$$(788 - 295).\tanh \sqrt{\frac{10.3768 \times 0.857534}{39.78 \times 0.0076795}} \times 0.25$$

$$\text{Or} \quad \dot{Q} = 710.32\ W \quad (27)$$

Comparison of results with CFD analysis.

- **Physical system and CFD modelling:** Dimensions of the pipe support and input parameters are taken as per respective problem statements (FIGS. 4 and 5). The type of element used for the simulation is a "tetrahedral element" throughout the computational domain. This choice of elements is advantageous with little distortion in mesh for complex geometries. The mesh is refined near the solid fluid interface to accurately capture the higher gradients in properties. A coarse mesh is otherwise used to save simulation time.

  This problem was studied numerically by solving the relevant governing equations (i.e., conservation of momentum equation, continuity equation and energy equation) for the given boundary conditions. In the study, the following assumptions were made:
  - Fluid flow over the pipe support is laminar and incompressible
  - Heat loss due to radiation is negligible

- ◦ Viscous dissipation is negligibly small.

  These assumptions are valid for the temperature and velocity ranges expected for the given problem.
- **Result:** The temperature distribution across the pipe shoe and trunnion supports is shown in FIGS. 6A and 6B, respectively. This distribution was obtained from CFD analysis for Problem Statements 1 and 2 with the same parameters.

  A quantitative comparison between the results obtained from the proposed method and CFD analysis is shown in FIGS. 7A and 7B, respectively. It demonstrates the temperature at the bottom of pipe supports at different air velocities. Red lines show the results from the proposed method and green lines show results from CFD analysis. It can be observed that the results obtained from the proposed method are in good agreement with the CFD simulation results.

For shoe support, variation in temperature with respect to CFD analysis is slightly on the higher end and due to additional inclusion of the shoe bottom plate in the CFD model that was not included during calculations using the proposed method. Due to consideration of the bottom plate during CFD analysis, more convection takes place and the temperature drop increases. As a result, with an increase in the air velocity, the difference between the results from both methods increases due to the increased convection at higher velocities. On the other hand, due to the absence of such approximation in the trunnion support, the deviation in temperature with respect to CFD analysis is lower and constant. Although the effect of convection through the bottom plate can be considered for more accurate results, this approximation makes the calculations easier and yields acceptable results.

Takeaway. An analytical method has been presented here to simulate the heat transfer across pipe supports. Analytical results were compared with CFD results and found to be in good agreement. Using this analytical approach, the temperature at the bottom of the pipe support can be calculated quickly and accurately, and an adequate support length can be established while complying with temperature limitations set by engineering consultant/client specifications. This eliminates additional cost and efforts required by FEA/CFD analysis using commercial software. HP

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Case study: Challenges in the selection of a helical baffled exchanger

In a greenfield project, a proprietary heat exchanger[a] has been found to be preferable to conventional shell-and-tube (S&T) exchangers to optimize the total lifecycle cost (i.e., capital, operating, installation and maintenance costs). In a crude preheat train, where the operating fluids are highly fouling and viscous, the unique geometry of the heat exchanger offers lower operating and maintenance costs, outweighing the higher capital cost and optimizing the total lifecycle costs of these exchangers. This article demonstrates that in a typical crude preheat train, a detailed comparative study between conventional S&T exchangers and the proprietary heat exchanger[a] is required before finalizing the type of exchangers. In this study, the sizes and performances of four exchangers in a crude preheat train were compared between conventional S&T exchangers with segmental baffles and the proprietary heat exchanger[a] for the same set of fluid properties and performance parameters.

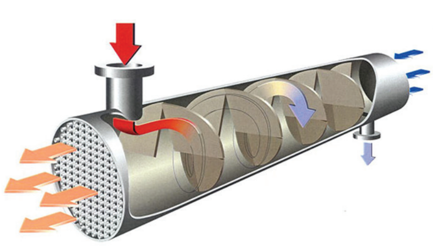

FIG. 1. Typical proprietary heat exchanger.

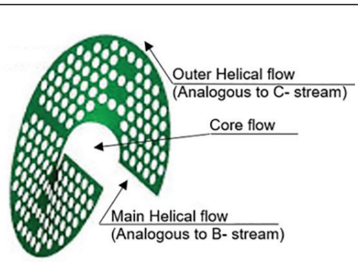

FIG. 2. Helical baffle flow regions.

Comparison. In the proprietary heat exchanger[a], quadrant-shaped baffle plates are positioned on the shell side at an angle to the tube axis, creating a helical flow pattern, as shown in **FIG. 1.**

The shell-side flow pattern is the main difference between a conventional S&T heat exchanger and the proprietary heat exchanger[a]. In a conventional S&T heat exchanger, a continuous change in flow direction is created by the baffles (resulting in ineffective use of pressure drop) and steady leakage of shell-side fluid exists through the baffle and shell clearance (resulting in wastage in effective heat transfer surface). The very nature of the proprietary heat exchanger[a] baffle construction creates a swirl to the shell-side fluid (thereby effectively utilizing the pressure drop in heat transfer) and continuously brings the fluid from the bundle periphery to the inside of the bundle (thereby effectively reducing the leakage stream).

Notable differences between segmental and helical baffles include:

1. While there are five flow paths in conventional S&T exchangers, there are three flow paths in the proprietary heat exchanger with helical baffle, as shown in **FIG. 2.** The flow paths are continuous and in parallel, unlike the alternate pattern of window and crossflow in the segmental baffles. Although the bulk of the flow (analogous to B-stream in a conventional S&T heat exchanger) follows the main helical path around the baffles, the outer helical flow (analogous to C-stream) and core flow are also important contributors to overall heat transfer.
2. The segmental baffle requires the correct selection of baffle cut and spacing to achieve a proportionate velocity ratio between the window and cross flow, to minimize loss of energy to changes in velocities in the window and crossflow regions. The helical baffle requires an ideal helical flow pattern to attain similar temperature profiles for fluid in each of the three flow paths.
3. The flow in a helical baffle will always have a longitudinal component.
4. There is no distinction between 45° and 90° layouts, or between 30° and 60° layouts for the helical baffle.

Ways in which the proprietary heat exchanger[a] surpasses conventional S&T exchangers are detailed in **TABLE 1.**

CASE STUDY

Four heat exchangers with fluid properties and performance parameters (listed in **TABLE 2**) in a crude preheat train were sized as conventional S&T exchangers with segmental baffles, as well as the proprietary heat exchanger using an incremental software program[b]. It was determined that, for the given fluid properties and performance parameters, conventional S&T exchangers with segmental baffles would have performed equally well without impacting the total lifecycle cost of these exchangers.

Results, discussion and lifecycle costs. For the crude/vacuum residue product (VR PDT) exchanger:

- The overall heat transfer rate was

only 4% higher for the helical baffle. The increase in overall heat transfer coefficient (U) due to improved heat transfer with the helical baffle was partly compensated by the use of a rotated square pitch with segmental baffle.

- A marginal improvement was seen in mean temperature difference (MTD) with the helical baffle due to the absence of leakage streams.
- The effective surface area was 6% lower for the helical baffle due to higher overall U and MTD, described above.
- The capital cost with the helical baffle will be much higher than with the segmental baffle—the reduction in cost of the proprietary heat exchanger due to the 6% reduction in surface area will be more than offset by the higher cost of the heat exchanger[a] design.

For the crude/VR PDT + quench exchanger-III:

- The overall heat transfer rate is almost same in both helical and segmental baffle designs. Any increase in overall U due to more effective heat transfer with the helical baffle was compensated by use of a rotated square pitch with the segmental baffle.
- The MTD is almost identical for both designs since the impact of the leakage stream in the segmental baffle design case in minimal.
- The effective surface area is the same in both helical and segmental baffle designs due to similar U and MTD values.
- The capital cost with the helical baffle will be significantly higher than the segmental baffle due to the higher cost of the heat exchanger[a] design.

For the crude/VR PDT + quench exchanger-II:

- The overall heat transfer rate is almost the same in both helical and segmental baffle designs. Any increase in overall U due to more effective heat transfer with the helical baffle was compensated by use of a rotated square pitch with the segmental baffle.
- Marginal improvement in MTD was achieved with the helical baffle due to the absence of leakage streams.
- The effective surface area is 4% greater for the segmental baffle due to the lower MTD described above.
- The capital cost with the helical baffle will be much higher than with the segmental baffle since the reduction in cost of the proprietary heat exchanger due to the 4% reduction in surface area will be more than offset by the higher cost of the heat exchanger design.

For the crude/VR PDT + quench exchanger-I:

- Marked improvement (~ 9%) was seen in the overall heat transfer for the segmental baffle, due to the change in exchanger arrangement to four shells in series with two tube passes rather than two shells in series and two in parallel with four tube passes in the case of the helical baffle design.
- Significant MTD improvement was seen in the segmental baffle design, where the negative impact of leakage streams is compensated by improvement in F factor to the increase in the number of shells in series from two to four.
- The increase on overall U and MTD resulted in a reduction in the effective surface area by 14% compared to the helical baffle design.
- The capital cost with the helical baffle will be significantly higher than with the segmental baffle due to a 14% increase in surface area and proprietary heat exchanger[a] design.

Capital and installation cost: It is evident that the capital and installation cost of the four services discussed here will be higher for the proprietary heat exchanger compared to conventional S&T exchangers with a segmental baffle.

Maintenance cost: The required time to reach the design fouling resistance value will be more with a helical baffle compared to segmental baffle exchangers due to higher shell-side velocity. Additionally, the cleaning frequency for the specified fouling factors will be less for helical baffle exchangers. Therefore, the maintenance cost will be lower for the proprietary heat exchanger compared to conventional S&T exchangers with a segmental baffle.

Operating cost: The shell-side calculated pressure drop for all four services with segmental baffles are

TABLE 1. Proprietary heat exchanger[a] scores over conventional S&T exchanger

Service/construction	Remarks
High viscosity fluid on shell side	Absence of baffle-to-shell bypassing and continuous drawing of fluid from the periphery to center of bundle
Reduced shell-side fouling	Due to near plug flow condition achieved throughout the bundle length
Smaller surface area	More heat transfer for the same allowable pressure drop due to better mixing of flow and no unnecessary change of direction of flow inside shell

TABLE 2. Fluid properties and performance parameters of four heat exchangers

Service	Crude/ VR PDT exchanger	Crude/VR PDT + quench exchanger-III	Crude/VR PDT + quench exchanger-II	Crude/VR PDT + quench exchanger-I
TEMA type[2]	AES	AES	AES	AES
Heat duty, MM Kcal/hr	4.4649	9.1234	9.4985	12.4146
Shell-side fluid	VR + slop	VR + slop + quench	VR + slop + quench	VR + slop + quench
Tube-side fluid	Crude	Crude	Crude	Crude
S/S viscosity (in/out), cP	10.5/16.5	4.9/7.81	4.36/6.42	2.94/4.36
S/S fouling factor, m^2-hr-°C/Kcal	0.002	0.002	0.002	0.002
T/S fouling factor, m^2-hr-°C/Kcal	0.000614	0.000819	0.000819	0.000819
S/S temp. (in/out), °C	251.1/227.1	279.5/249.4	309.8/276.3	351.5/309.8
T/S temp. (in/out), °C	201.4/212.2	223.5/244.7	246.4/267.8	266.7/313.2
Allowable DP (S/S), kg/cm^2	4.1	4.9	2.8	1.5

TABLE 3. Size and performance data for both helical and single segmental baffle designs

Parameters	Crude/VR PDT exchanger		Crude/VR PDT + quench exchanger-III		Crude/VR PDT + quench xchanger-II		Crude/VR PDT + quench exchanger-I	
Baffle type	Helical	Single segment	Helical	Single segment	Helical	Single segment	Helical	Single segment
Shell ID, mm	1,250	1,180	1,320	1,280	1,480	1,480	1,400	1,380
Tube length, mm	6,000	7,000	6,000	6,500	8,000	8,500	8,000	7,315
No. of shells, S / P	2 × 1	2 × 1	4 × 1	4 × 1	2 × 1	2 × 1	2 × 2	4 × 1
No. of passes	2	2	2	2	4	4	4	2
Tube OD, mm	25	25	25	25	25	25	25	25
Tube layout, degrees	90	45	90	45	90	45	90	45
Baffle pitch, mm[1]	205	250	260	275	255	300	205	446
Baffle cut, %	Note 2	25.6	Note 2	25	Note 2	22.5	Note 2	25.4
Surface area, m^2	808.6	858.5	1,775.6	1,773.4	1,521.6	1,578.7	2,763.6	2,433
MTD, °C	31.6	31.4	30	29.9	34.8	34.5	28.6	29.8
Calculated DP (S/S), kg/cm^2	2.2	1.8	4.9	4.7	2.8	1.6	1.5	1.5
Avg. velocity, m/sec	1.5	0.74	1.8	0.97	1.5	0.63	0.87	0.51
Heat transfer co-efficient (S/S), m^2-hr-°C/Kcal	519	447	643	645	626	535	497	593.5
Overall U, m^2-hr-°C/Kcal	174.7	167.5	171.3	176.8	179.4	176.3	157.1	171.7
Thermal resistance: shell, % of total	33.6	37.5	27.2	27.4	29.5	32.9	32.6	28.9
Thermal resistance: fouling, % of total	48.5	46.7	53.3	53.9	56.3	56.7	49.4	52.4

[1] Baffle pitch assumed for proprietary heat exchanger[a]   [2] Baffle cut is not applicable for proprietary heat exchanger[a]

the same—if not lower in some cases—than those with helical baffles. Therefore, the operating cost (which is related to the pressure drop) of these exchangers with a segmental baffle will be same, if not lower, compared to those with a helical baffle.

Total lifecycle cost: As noted here, all components of the total lifecycle cost—excluding maintenance cost—are lower for the conventional S&T exchanger with segmental baffle option compared to the proprietary heat exchanger option for the four services considered here. Since a lower maintenance cost will be more than compensated by the higher capital, installation and operating costs of the proprietary heat exchanger, compared to a conventional S&T exchanger, the proprietary heat exchanger[a] option will be unviable from a total lifecycle cost perspective.

Total lifecycle cost optimization: Since the average shell-side velocity with a helical baffle is about 1.5–2 times the velocity with a segmental baffle, the proprietary heat exchanger will have a lower propensity for shell-side fouling. With a fouling factor much lower than 0.002 Kcal/m^2-hr-°C, which is more appropriate for a segmental baffled exchanger, the size of the proprietary heat exchanger can be considered. Since the overall size—and, therefore, the capital and installation costs—are dependent on additional surface area required to overcome fouling (all four sizes are controlled by fouling, as explained above), a lower fouling factor would have resulted in a much lower surface area for the proprietary heat exchanger. This would have resulted in a lower capital cost, making the proprietary heat exchanger[a] option viable in terms of the overall lifecycle cost.

Takeaway.

While selecting the proprietary heat exchanger, the inherent advantage of higher shell-side velocity with a helical baffle must be considered while specifying the shell-side fouling factor. A conservative approach of using the same fouling factor as a conventional segmental baffled exchanger can make the proprietary heat exchanger option unviable.

While selecting an alternate option of a conventional S&T exchanger with a segmental baffle, careful shell arrangement selection is vital for multiple shell exchangers, tube and baffle geometries to optimize the use of allowable pressure drop and make the exchanger more compact.

The total lifecycle cost, rather than only CAPEX, must be considered by owner-operators and contractors during front-end engineering and design (FEED) when selecting the best heat exchanger option for a particular service. HP

NOTES

[a] Lummus Technology's HELIXCHANGER™
[b] HTRI Xist

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Achieve greater reliability with a redesigned instrument manifold

Across industries, ranging from chemical upstream oil and gas to food and beverage applications, the manifold has proven to be a solid and hard-working instrument that receives little attention until it stops performing as designed. The ability to equalize or shut off pressure is an essential part of the efficiency required to operate a plant. However, the traditional manifold design has remained the same, with reliability issues tied to viscous applications that contain abrasive materials or grit, which can impact the operation of the manifold (FIG. 1).

Improving an existing design requires a deep understanding of what that piece of equipment does and why it is important. Manifolds are put in place to efficiently and safely isolate the pressure sensor so maintenance can be performed. Systems and processes that deal with abrasive or dangerous materials require that the manifolds operate reliably and minimize liability.

The elements of the manifold's standard design that cause the primary reliability issue are those that can cause or allow process fluid to reach and damage the transmitter. When process fluid damages the manifold, even before it reaches the transmitter, it can distort the accuracy of the readings, which can significantly impact the process in the facility.

A pressure-lock design. It is easier to securely position the manifold with a two-piece stem design and a non-rotating valve tip (FIG. 2) than with a standard manifold. Creating a more secure seat and providing closure with minimal wear ensures equipment lasts longer and is not affected by process fluid. In challenging processes—where sand and grit in the flow can cause unexpected pressure changes or damage to the pressure sensor—an operator may need to quickly close a valve manually. Doing so with a two-piece design will make it easier to

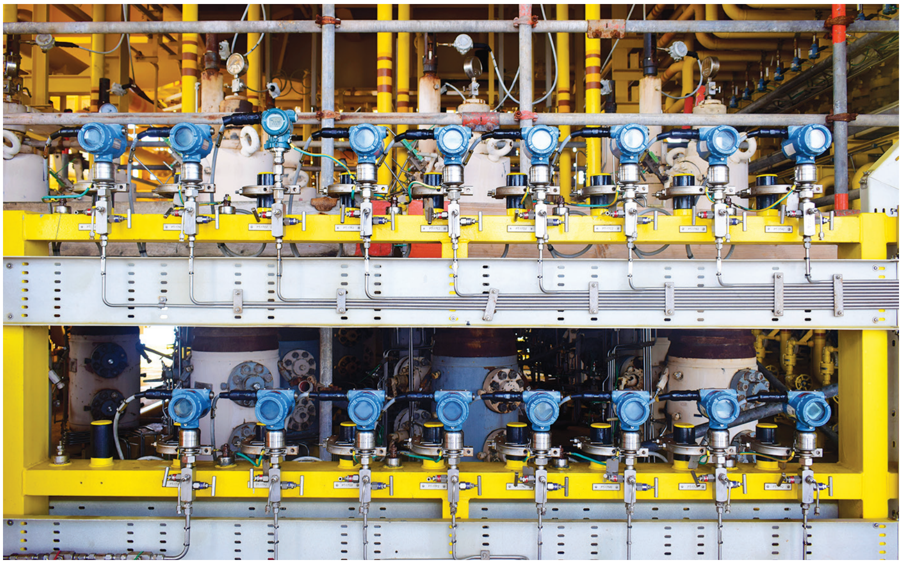

FIG. 1. Instrument manifolds provide isolation, venting and equalization for various types of pressure measurement applications.

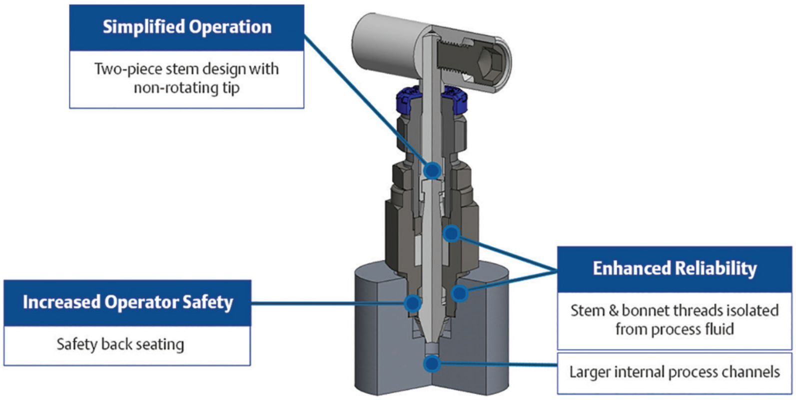

FIG. 2. Unlike standard manifold needle valves, enhanced manifolds with a two-piece stem design offer simplified operation, enhanced reliability and increased operating safety.

turn the handle, saving time as well as reducing the risk of injury to the operator gripping the handle. An added benefit is that a manually closable manifold creates a mechanical barrier, keeping process fluid where it should be.

An instrument manifold designed with a wider bore channel makes clogging less likely, as small sedimentary material or viscous fluids can more easily flow through the manifold. This makes it more suitable for applications that can have sand, grit or even sludge in the process flow, such as found in fracking, mining, or paper and pulp industries.

Consider, too, that an instrument manifold designed with a wider bore channel makes clogging less likely, as small sedimentary material or viscous fluids can more easily flow through the manifold. This makes it more suitable for applications that can have sand, grit or even sludge in the process flow, such as found in fracking, mining, or paper and pulp industries.

The manifold at work. In a plant environment where dozens of manifolds may be operating at one time, it is up to the process engineer to control valves in a pressure measurement application by way of manifolds. The easier the manifold is to use, the more efficient the process engineer can be, and the lower the risk of process upsets from delayed valve opening or closing. Reducing the risk of process fluid leaks using better-designed equipment (**FIG. 3**) ensures the safety and longevity of the equipment.

Opening and closing a valve manually is not necessarily a daily occurrence in a plant, especially when pressure can be read digitally; however, having a user-friendly device is preferred when it does need to be done. Being able to stop the process fluid to conduct routine scheduled transmitter maintenance is one instance when manual operations are required.

Applications with particularly challenging material flow may benefit from the wider bore channel. For example, hot tar, maple syrup or even peanut butter can all quickly build up and clog a manifold, making it harder to get correct measurements. In extreme cases, if the buildup becomes too great, it can cause a blowout or manifold failure. This can mean that pressure behind the manifold becomes too great because the material cannot flow through at the appropriate speed, plugging the line and causing pressure to build. Avoiding these dangers is a benefit of the wider bore channel.

Manifolds in the field. An example of utilizing a pressure-lock design is found in chemical plants with reactions that can quickly spike in pressure. These processes are monitored closely because fast reactions or potentially valuable or hazardous processes must not, under any circumstances, be allowed to leak into the ambient environment. Therefore, a valve must be sealed tightly after maintenance—a manifold that is easier to manipulate ensures the valve is sealed more securely.

In oil field applications, it is imperative that manifolds operate correctly, ensuring that operators can read pressure spikes from a well. A missed, delayed or inaccurate reading can mean that abrasive process fluids can reach the sensor, thereby damaging it. Undetected spikes both in chemical and oil field applications spell safety concerns with potentially catastrophic consequences.

If a manifold requires special tools or is difficult to operate, it makes the process more difficult and can lead to repetitive strain injuries for the operator who must continually adjust the manifold. Especially in the field, out in the elements and where actions sometimes must be made quickly, avoiding injury and keeping operators safe are important considerations when sourcing a manifold.

The evidence is clear. A better-designed manifold can help processes run more smoothly and make the life of the operator easier. The two-piece stem design with a non-rotating valve tip makes operations run more smoothly, while the adjustable packing nut simplifies maintenance and allows for easier open and close operations. The added benefit of having the stem and bonnet threads isolated from the process fluid minimizes the potential for corrosion.

Meeting industry requirements and providing a better device, especially for highly corrosive, gritty or sludge-like materials in the process flow, ensures greater efficiency, reliability and longevity for equipment that requires manifolds. HP

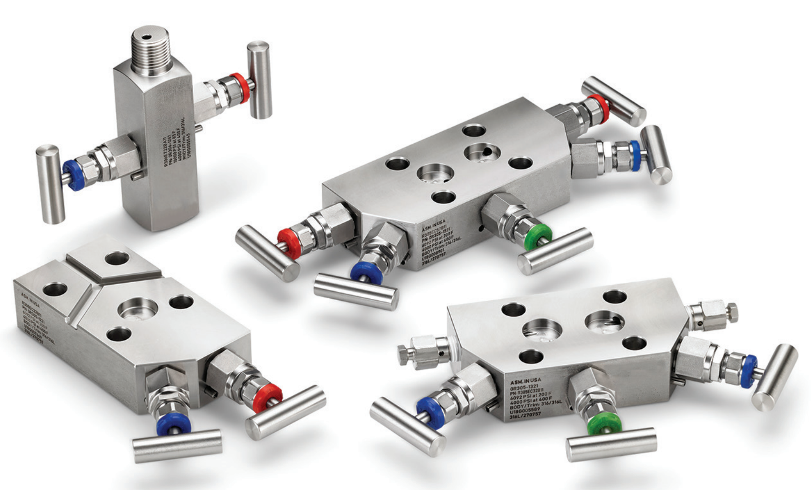

FIG. 3. Enhanced manifolds enable compatibility with a variety of pressure instrumentation.

BRANDON HASCHKE is a Product Manager focusing on instrument manifolds for Emerson's Rosemount instrumentation business. He now concentrates on helping operators in a variety of industries get the most out of their pressure measurement points. Mr. Haschke holds degrees in chemistry and chemical engineering from the University of Minnesota.

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Are hot pipes and equipment surfaces ignition sources in the HPI?

Most international and national standards used in the hydrocarbon processing industry (HPI) for conducting hazardous area classification do not provide any guidance or recommendation for non-electrical equipment, such as hot pipe and equipment surfaces, that can act as ignition sources. These standards are written specifically for electrical equipment as the source of ignition for hazardous area classification purposes. Accordingly, HPI engineers are also conducting this exercise for the selection of electrical equipment in hazardous areas and other non-electrical ignition sources that are not reported in the classification process.

Approach. Some of the international and national standards used for the area classification exercise and their position with respect to non-electrical ignition sources are discussed here.

IS 5572[1] does not apply to ignition sources other than those associated with electrical apparatus. In fact, its title and scope clearly show that it is applicable only for electrical apparatus, clarifying that in any plant installation, irrespective of size, there may be numerous sources of ignition apart from those associated with electrical apparatus. It goes on to say that these are outside its scope and additional precaution may be necessary to ensure safety in this aspect.

OISD-STD-113[2] deals with the classification of areas for electrical installations at hydrocarbon processing and handling facilities and, like IS 5572[1], the guidance on non-electrical ignition sources is outside the scope of this standard, as well.

API 500[3] and API 505[4] deal with the classification of locations for electrical installations in petroleum facilities. The suitability of locations for non-electrical equipment is beyond the scope of both API 500 and API 505.

NFPA 497[5] is a recommended practice for the classification of flammable liquids, gases or vapors and of hazardous (classified) locations for electrical installations in chemical process areas. It does not provide any guidance about the presence of other ignition sources, such as hot pipe and equipment surfaces and other non-electrical equipment.

IEC 60079.10.1[6] is concerned with the classification of areas where flammable gas or vapor hazard may arise. However, it does not provide guidance about the presence of other ignition sources, such as hot pipe and equipment surfaces and other non-electrical equipment.

AS/NZS 60079.10.1:2009[7] does acknowledge that flames, incandescent material, hot surfaces and mechanical impact sparks are sources of ignition. However, no guidance or recommendations on non-electrical equipment are provided in this standard.

The 3rd edition of EI 15[8] provided guidance and recommendations on non-electrical sources of ignition. However, in its current 4th edition, those guidance and recommendations have been removed and incorporated in a different part of model code of safe practices that is not normally referred by engineers while conducting hazardous area classification.

Based on the study of international and national standards here, it is clear that most of them do not provide any guidance or recommendations for non-electrical ignition sources.

Hazardous area classification. The concept of hazardous area classification was originally developed for the selection and location of fixed electrical equipment for use where flammable atmosphere was anticipated. However, inappropriately selected electrical equipment are not the only source of potential ignition—items such as fired surfaces, hot surfaces, frictional sparks, etc., are also potential ignition sources.

Equipment designed for controlled combustion are inevitably hot and hazardous area classification cannot be sensibly applied close to them. Such equipment normally encountered in a plant include fired heaters, flares, gas turbines and internal combustion engines. The location of such equipment should be determined during plant layout. A good practice is to locate such equipment as far as practical outside the hazardous area.

Risk should be considered when hot pipe and equipment are installed in hazardous area. Appropriate selection of equipment depends on the zone, gas group and auto ignition temperature (AIT) of the fluids present. The AIT is a particular concern as fluids heated to above the AIT can ignite when mixed with air without any ignition source.

Fluid properties. Some of the important fluid properties that are required for hazardous area classification are flashpoint, AIT, lower flammability limit (LFL) and upper flammability limit (UFL). Flashpoint and AIT are two entirely different properties of a fluid. The flashpoint of a liquid is the minimum temperature at which the vapor generated can form a flammable atmosphere

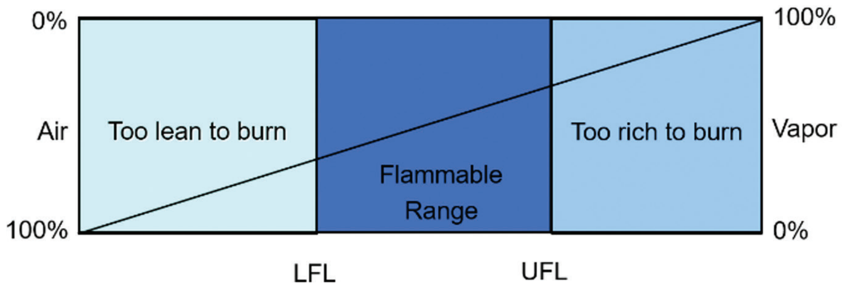

FIG. 1. Stoichiometric concentration.

TABLE 1. AIT of typical flammable fluids handled in the HPI

Fluid	AIT, °C
Benzene	498
n-butane	288
Carbon monoxide	609
Ethane	472
Ethylene	400
Gasoline	280
Hydrogen	500
Hydrogen sulfide	260
n-heptane	204
n-hexane	225
Kerosene	210
Liquified petroleum gas	405
Methane	600
Methanol	385
Naphtha	288
n-nonane	205
n-octane	206
n-pentane	243
Propane	450
Propylene	460
Toluene	480
Xylene	464

in air. Ignition does not take place at the flashpoint unless there is an ignition source. If the liquid is heated further, at some point (AIT) it will ignite without the application of an external ignition source. This temperature is higher than the flashpoint. AIT is the lowest temperature at which a flammable gas or vapor ignites by itself. AIT varies widely for hydrocarbon, from 600°C for methane (CH_4) down to 204°C for heptane (C_7H_{16}). Values for typical flammable fluids are given in NFPA 497[5] and ISO/IEC 80079-20-1[9]. Common fluids handled in the HPI are listed in **TABLE 1.**

Auto ignition of a fluid depends upon numerous factors, including:

- How near a vapor/air concentration is to the stoichiometric concentration
- The temperature of surfaces in contact with the fluid
- The contact time between the surface and the fluid
- The area of the hot surface.

The stoichiometric concentration is the concentration of vapor between the LFL and UFL of that fluid in the air, also known as flammable range, as depicted in **FIG. 1.** The LFL of a vapor is the concentration of that vapor in air below which the vapor concentration is too lean to burn. The UFL of a vapor is the concentration of that vapor in air above which the vapor concentration is too rich to burn. So, for ignition to take place, the concentration of vapor should be between the LFL and UFL. Values of LFL and UFL for typical flammable fluids are given in NFPA 497[5] and ISO/IEC 80079-20-1[9].

It is difficult to raise the temperature of a gas or vapor to above the AIT under open air ventilation conditions encountered in the HPI. It is still a good practice to avoid very hot surfaces in hazardous areas, even with open ventilation. In areas with less ventilation, it is easier for a hot surface to heat a gas or vapor release to above the AIT and, therefore, equipment surfaces should not exceed the AIT of any release.

For the HPI, based on the lowest AIT encountered (that of heptane), a limit of 200°C is recommended. This means that hot pipe and equipment surfaces up to 200°C cannot ignite any hydrocarbon in open-air ventilation conditions. Also, it is a normal practice in the industry to insulate hot pipe and equipment surfaces to prevent energy loss and for personnel protection. This ensures that the hot surface is not readily available for contact with the flammable atmosphere, thereby mitigating the risk of ignition. This mitigation considered during the design phase assumes that the plant will be timely inspected and well-maintained to ensure that the insulation is continuous and intact.

This discussion justifies disregarding hot pipes and equipment as ignition sources during hazardous area classification. However, this is not captured in any document—as standards are also generally silent on this aspect, this becomes a grey area for engineers designing, maintaining and operating hydrocarbon plants. It is recommended that this aspect is clarified in the hazardous area classification documents so that no ambiguity exists in the later stages of the plant life.

Takeaways. All sources of ignition must be controlled where flammable atmospheres may form, and hazardous area classification is a good basis for deciding what equipment may be used at any location. While the principal objective of hazardous area classification remains the classification of areas for the selection of an appropriate type of electrical apparatus, the design documentation for hazardous area classification should provide additional guidance to aid the location and control of non-electrical sources of ignition. It should include justification for not considering the hot surfaces in the plant, such as equipment and pipe surfaces, as sources of ignition in the HPI. HP

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Upkeep down the pipe: Plant shutdowns and water treatment

The maintenance checklist for a plant shutdown can sometimes look intimidating and, in the face of a short timeline, unnerving. For aging facilities, every year the maintenance checklist becomes more extensive and completion becomes more difficult. As staff faces greater challenges, bringing in specialists to help complete certain tasks is almost imperative. All facilities require water to operate, and a water treatment specialist should be one of the first outside contractors to be introduced to the shutdown team (FIG. 1).

Inevitably, once maintenance has begun, other issues and concerns can arise. When it comes to the water system (steam boilers, cooling towers, pre-treatment systems, etc.), these issues must always be addressed immediately to maintain production and avoid costly damage to equipment. Waiting to call a water treatment specialist only during emergencies can be expensive. A water treatment-related emergency can raise the cost of service calls by 2–3 times, garner a premium for parts and cause unscheduled downtime. Shutdown procedures should help prevent these inflated expenses. Having a scheduled water treatment specialist integrated into the plant shutdown plan shortens the maintenance checklist so internal staff can focus on other things. It also promotes faster and safer diagnoses and resolutions that mitigate costly unscheduled downtime.

Water system issues are not always what they seem.

In the face of increased corrosion, heavy deposition, scale, microbiological fouling and system leaks, operators often ask, "If a system was working fine for years, what changed?" Reasons can include:

1. Higher dissolved oxygen content in the treated water
2. Changes in feedwater chemistry
3. Mineral or microbiological content and concentration
4. Aging or faulty equipment (seals, pipes and valves)
5. Greater demands on the system due to weather or increased production.

These causes start affecting the system through biofilm growth, corrosion, scale and accumulating deposits. Generally controlled by water treatment products, changes in feedwater or system conditions can throw off the ratios of treatment products, rendering them less effective. Not only does this affect system health, but it also greatly reduces system efficiency.

Why would feedwater quality change? Climate change has proven to have drastic effects on water quality, particularly in times of reduced snowpack. Upstream issues, such as industrial monoculture farming and mining/quarrying can increase sediment levels, mineral constituents and chemical content. Changes in water sources, such as moving from surface water to deep aquifers, will result in drastic changes in water chemistry and quality. Lighter permitting standards and loosened wastewater runoff regulations can also impact the water table for surrounding communities.

Keeping on schedule.

Merely relying on accrued work orders to determine the maintenance checklist is an effective way to discover issues that throw the schedule off by days, even weeks. Thorough inspections by outside professionals several weeks before a scheduled production break can introduce new eyes for inspection purposes, help in planning the full scope of the shutdown, inform proper part orders and reduce the total downtime.

FIG. 1. The water treatment team focuses on details in every component of the water system.

One major cause of delays that extend downtime is waiting for vital replacement parts. Generally, these are ordered beforehand; however, not all issues can be anticipated. Having a water treatment specialist involved in the shutdown might lead to a faster solution. They have the skills and knowledge to provide support to operators and contractors or perform much of the maintenance and repair work on water treatment equipment.

For example, the technician can perform maintenance on dosing systems, from replacing pump diaphragms and liquid ends to tubing and injection quills. Upgrading controllers, or installing, programming and commissioning new ones can take excessive time when done by untrained personnel. This also applies to replacing ion exchange and other pre-treatment equipment controllers, heads and media, as well as programming and commissioning them. If reverse osmosis or nanofiltration membranes require cleaning or replacement, the technician can quickly and efficiently complete that task. By having the water treatment specialist perform this work, operators are free for other shutdown tasks while knowing that the job is performed properly.

FIG. 2. Inspections require an experienced eye for traces of possible problems.

FIG. 3. A water treatment specialist personally inspecting the steam drum of a boiler.

Water treatment: Planning for a shutdown. The moment the water leaves the ground to the moment it is released to the municipal sewer lines is the space where water treatment operates. After integrating with the internal team to conduct the inspection and prep a shutdown action list, water specialists should collaborate with supervisors to create a maintenance schedule that coincides with their timeline. This includes meeting with scaffolding, plumbing and electrical contractors to coordinate workflows, as well as reviewing safety permits to create lockout procedures and present a workplace safety plan.

With a team of as many as five technicians, the water treatment staff will check every inch of the system using a series of inspection and testing methods, applying years of experience in the field to identify potential issues (**FIG. 2**).

Treatment product pumps, injectors, tubing and valves can succumb to time, particularly if they are improperly inspected and maintained. Containment tanks and storage tanks can display signs of corrosion, blockages, leaks and other physical damage that can cause danger to operators, waste products, increase costs and detrimentally affect the dosing rates, leading to insufficient treatment. Probe and chemical injection piping racks must be inspected and cleaned.

Often, valves that may have never been turned or resealed slowly corrode without showing external signs of degradation. Corrosion monitoring stations can detect some of these issues, but they also require inspection for proper operation. Conversely, deposits can form that will prevent the proper seating of valves, slow velocity through the piping and reduce heat transfer efficiency in heat exchangers. Clean-in-place (CIP) systems can remedy some of these issues by using a mix of chemicals, heat and water to clean process pipes, vessels and machinery. CIP systems reduce or eliminate the need to dismantle parts of the system.

What is discovered when checking and maintaining filtration systems may impact what must be done during a shutdown. Changes in feedwater quality can cause the media in carbon, multi-media, softeners or other filtering units and the membranes in reverse osmosis systems to foul more quickly, leading to reduced efficiency, production rates and pressures. If the water quality has changed dramatically, resetting the proper maintenance schedule for these units requires testing and monitoring. Ion exchange units may need a deep-cycle clean or require resin replacement to ensure throughput and exchange capacity are maintained.

Case study: Evidence in the boiler. An oil and gas facility in Alberta planned a shutdown with all the proper checklists and schedules in place. Managers included a team of special technicians from the site's water treatment servicer to assist with the maintenance, cleaning, refurbishing and installation of equipment related to the boiler system.

Although a specialist was invited to do a preliminary inspection, a full assessment was impossible until the system shutdown had fully drained the boiler and pipes. Once that was accomplished, technicians were able to get inside the vessels.

A close inspection with the naked eye is vital, as shown in **FIG. 3.** If the specialist's participation is consistent, pictures are compared to the previous shutdown, as well as samples that provide a clearer idea of the scope of any issue.

Climbing inside the deaerator storage tank of the boiler, technicians found dark red particulates at the bottom, indicating excessive iron corrosion of a system component. A specially formulated polymer blend was dosed into system water to sequester dissolved solids and iron, holding them in suspension to be flushed from the boiler during regular blowdowns. In this case, the amount of particulate matter exceeded the amount the treatment polymers being used were intended to treat. This resulted in particulates deposited in the condensate tank to be carried with the feed water to the boiler, as well.

If left unaddressed, deposits would collect on the boiler internals, diminishing heat transfer efficiency and promoting under-deposit corrosion. Corrosion can cause pipe and tube walls to succumb to system pressure, potentially leading to a sudden rupture. Heated water or steam can escape, causing expensive leaking or flooding. In the case of steam boilers, catastrophic tube failures can lead to explosions and, if staff is nearby, serious injuries.

Technicians walked the entire system, testing and searching for signs of corrosion. After a thorough examination, they discovered the problem was coming from a corroding valve to a bypass line. These lines are used for maintenance if issues in the main feedwater line system require other pre-treatment equipment to be temporarily bypassed. Bypass lines commonly remain flooded, and this valve had stayed in a single position for several years without showing any outer signs of corrosion. On the bypass side of the valve, the water lacked corrosion-prevention treatment, causing the valve to corrode virtually imperceptibly from the bypass side.

The team diagnosed the causes, replaced the valve and repaired any further corrosion around the area. They then thoroughly checked the other valves and recommended maintenance procedures and schedules in the final report to eliminate additional occurrences.

Takeaway. In many industries, scheduled plant shutdowns are unavoidable and necessary. Some maintenance activities are just too dangerous or impossible to perform during production. For water systems, some tasks can be performed without water flow or draining the system, but not many.

A proper shutdown procedure for water treatment requires an experienced supervising technician who knows how to identify problems, anticipate issues and plan a project schedule that integrates well with the other maintenance teams. They can work closely with internal water specialists or maintenance managers to ensure a smooth shutdown with no extensions or delays. HP

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GAS PROCESSING & LNG
JAN/FEB 2022

EDITORIAL COMMENT

Lee Nichols, Editor-in-Chief/ Associate Publisher

One thing that is constant is change. Every person, company and industry will experience this throughout their existence. Like all these entities, *Gas Processing & LNG* is not immune to change.

In late 2021, the publication's Editor-in-Chief, Adrienne Blume, moved to a different position at a scientific magazine outside of the energy industry. Ms. Blume was the publication's first Editor-in-Chief and was instrumental in its success over the past several years. Those responsibilities have now passed to me.

For readers of *Hydrocarbon Processing*, my name will be familiar. For the past several years, I have been Editor-in-Chief/ Associate Publisher of the centennial publication focused on the advances in refining and petrochemical technologies. My new responsibilities will continue to keep the high editorial reputation of not only *Hydrocarbon Processing* but that of *Gas Processing & LNG* and our newest hydrogen publication *H2Tech*. These three global publications reflect the advancements in technological ingenuity within the oil and gas and energy sectors and celebrate the latest technologies optimizing operations and producing products that are advancing the standard of living for billions of people around the world.

It is our mission to publish the latest technologies, services and products to the global oil and gas and energy industries. We will continue to fulfill this mission because these industries are needed to ensure the advancement of society and provide products and power that the world demands. **GP**

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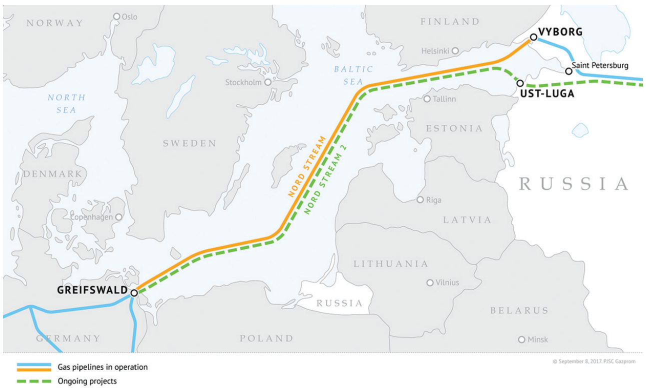
Cover Image: A view of SIBUR's Yuzhno-Balyksky gas processing plant, which has an associated petroleum gas processing capacity of 3 Bm^{3}y. Photo courtesy of SIBUR.



Nord Stream 2 ready for commissioning

Russia's Gazprom announced in late December that the $11-B Nord Stream 2 pipeline is ready to start sending natural gas to Germany. The natural gas pipeline, which will span the length of the Baltic Sea, will run approximately 1,200 km from Western Russia to Greifswald, Germany. Once commissioned, the pipeline will send 55-Bm^{3}y of natural gas to Western Europe. However, Russia's actions against Ukraine may delay the start of natural gas flows to Europe—at the time of this publication, Germany regulators have yet to sign off on the project and announced they may not approve the pipeline until 2H 2022. Once approved, Nord Stream 2 will help satisfy natural gas demand in Western Europe.

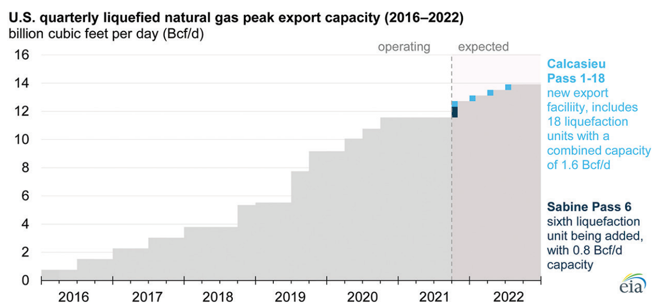
New Fortress Energy to develop an energy hub in western Africa

New Fortress Energy (NFE) signed a Memorandum of Understanding (MoU) with the government of Mauritania to develop an energy hub in the country. The hub will utilize the country's offshore natural gas reserves to develop natural gas, power, LNG and blue ammonia projects.

According to NFE, the company will deploy its Fast LNG liquefaction technology to produce LNG for local gas and power markets, as well as international exports. NFE will also supply natural gas to the 180-MW Somolec power plant and a new 120-MW combined-cycle power plant that is planned to be built. The MoU will also support the development of blue ammonia projects in the country.

Shell to supply CO_2 capture technology for Calpine's energy center

Shell Catalysts & Technologies will provide its Cansolv CO_2 technology for a natural gas combined-cycle power plant at Calpine's Deer Park Energy Center in Deer Park, Texas (U.S.). The post-combustion CO_2 capture facility will be designed to capture approximately 5 MMtpy of CO_2. Shell, along with TechnipEnergies, will also provide FEED work for the project. The FEED study is expected to be completed in early 2023. The Deer Park Energy Center supplies a baseload capacity of more than 1,100 MW to the Texas grid system.



U.S. to become largest LNG exporter this year

According to data by the U.S. Energy Information Administration (EIA), the U.S. became the third-largest LNG exporter in 2020 and will become the world's largest LNG exporter once additional trains at Sabine Pass and Calcasieu Pass begin operations this year.

By the end of 1Q, Cheniere plans to start Train 6 at its Sabine Pass LNG site in Louisiana. Once operational, the 5-MMtpy Train 6 will increase total LNG production capacity to 30 MMtpy.

Venture Global also plans to start exporting LNG out of its Calcasieu Pass LNG facility in the first half of this year. The $10-B project, located in Louisiana, consists of 18 liquefaction trains.

The completion of these two projects will increase U.S. LNG export capacity to nearly 14 Bft^{3}y. Additional U.S. LNG projects are also being developed. Gulf Energy Information's Global Energy Infrastructure database is tracking more than 350 MMtpy of active LNG liquefaction projects in the U.S. However, many of these projects have not reached final investment decision (FID). According to the International Gas Union, the U.S. accounts for nearly 40% of pre-FID capacity globally.

Mitsui and Vopak to own/operate largest FSRU

Mitsui and Vopak have signed an agreement to jointly own and operate the FSRU *Challenger*, which upon completion, will be the largest FSRU in operation. The FSRU, which will be located approximately 25 km offshore Hong Kong Island, will have a storage capacity of 263,000 m^3 and a regasification capacity of 800 MMft^{3}d. The vessel is scheduled to begin operations in mid-2022. Once operational, the FSRU will provide natural gas to domestic power plants.

JERA retires four LNG power plants, seeks to build three new ones

JERA, Japan's largest power generator, announced that it has retired four old LNG-fired power plants at its Anegasaki power station near Tokyo, Japan. The retirement of the older plants will make way for three new, less-polluting LNG plants. JERA said that the new LNG units will start operations in 2023.

Höegh LNG to provide FSRU for São Paulo

To help satisfy Brazilian natural gas demand, Höegh LNG has signed a charter deal with Brazilian gas trading and distribution company Compass Gas and Energia. Höegh LNG plans to install the 170,000-m^3 *Höegh Gannet* floating storage and regasification unit (FSRU) at Compass Gas and Energia's subsidiary—Terminal de Regaseificacao de GNL de São Paulo—in São Paulo. The FSRU is part of a planned 4-MMtpy import terminal at the site.

The new FSRU, one of five LNG imports planned for Sao Paulo, is scheduled to begin operations in late 2022/early 2023.

Venice Energy to build LNG facility at Port Adelaide

Venice Energy has received approval from the Australian government to build an LNG import facility in Port Adelaide, South Australia. The more than $180-MM Outer Harbor LNG project plans to install an FSRU to supply natural gas to South Australia. Construction is scheduled to begin in mid-2021, with operations to commence in late 2023/early 2024. This LNG import project is one of several LNG import projects being developed in Australia. Most LNG import projects in Australia are in the states of New South Wales and Victoria.

Petronas awards FEED contract for Sabah LNG project

Petronas awarded two front-end engineering design (FEED) contracts for the 2-MMtpy Sabah LNG project. The FEED contracts were awarded to Saipem SpA and the JGC Corp.-Samsung Heavy Industries JV. The winner of this dual-FEED contract competition will be awarded the engineering, procurement, construction and commissioning contract. The new near-shore LNG terminal will be built near the Sipitang Oil and Gas Industrial Park in Sipitang, Sabah, Malaysia. It will receive natural gas feedstock from the offshore Sabah gas fields. If built, the Sabah LNG facility will begin operations by 2027, increasing Malaysia's total LNG production to approximately 4.7 MMtpy.

Novatek's Arctic LNG 2 project secures financing

Novatek announced that its nearly 19-MMtpy Arctic LNG 2 project has secured financing. The $21-B, three-train LNG export facility is being developed on the Gydan Peninsula in Northern Russia. FID on the project was reached in 2019. The first train is scheduled to begin operations in 2023. All trains are expected to be operational by 2027. The LNG facility will provide natural gas to Asia and Europe.

Maire Tecnimont awarded engineering works by Eni for a carbon capture plant

Maire Tecnimont announced that its subsidiary, NextChem, has been awarded an engineering contract from Eni for a carbon capture plant at the natural gas plant of Casalborsetti, Italy. The plant's goal is to separate carbon dioxide (CO_2) from emissions from the natural gas plant's turbo compressor, purify and compress it. The new plant is designed to capture 25,000 tpy of CO_2.

New LNG terminal to be built in Karnataka, India

LNG Alliance, the Government of Karnataka and New Mangalore Port Trust plan to build a large-scale LNG import terminal in Karnataka, India. The facility's first phase includes the construction of a 4-MMtpy LNG import facility, with the potential to double to 8 MMtpy, if needed. The LNG import terminal plans to utilize a $290-MM FSRU to import natural gas that will be used to satisfy increasing demand in the state of Karnataka. Additional infrastructure will include an LNG truck loading facility and an LNG bunkering facility. This additional infrastructure will enable the facility to provide LNG supplies to the state's transportation sector and marine vessels.

QatarEnergy awards major contract for North Field expansion project

QatarEnergy awarded an EPCI contract to McDermott to help expand North Field natural gas operations in Qatar. According to McDermott, the contract includes 13 normally unmanned wellhead topsides, in addition to various connecting pipelines and the shore approaches for the North Field East pipelines, beach valve stations and buildings. The jackets and the pipelines for the North Field South project will be subject to a separate tender, which is expected to be awarded in the first half of this year.

The expansion project will enable Qatar to increase domestic LNG liquefaction capacity from 77 MMtpy to 126 MMtpy by 2030. This increase in capacity will be possible through additional natural gas feedstock from the North Field East and North Field South expansion projects. First LNG is expected to be produced in 2025.

Siemens awarded major power contract in Brazil

Siemens has been awarded a turnkey construction contract for a combined-cycle power plant in Brazil. The UTE GNA 2 power plant is part of the integrated LNG-to-power project GNA 2—GNA 1 began commercial operations in late 3Q 2021. The power plant will be in the Port of Açu, approximately 290 km from Rio de Janeiro in the Brazilian state of São Paulo. The more than $1.13-B project will add an additional 1.7 GW of power, and includes the construction of a second combined-cycle power plant, an LNG regasification terminal (FSRU) and transmission infrastructure.

Global gas and LNG: Six trends to watch in 2022

In the report "Six things to watch for in 2022," Wood Mackenzie provided insights into six major trends affecting the global natural gas and LNG sectors in 2022. One, the start of operations of Russia's Nord Stream 2 pipeline could lower natural gas prices for Western Europe; however, tensions between the regions could hamper pipeline flows. Two, oil-indexation levels are expected to rise, potentially reaching 12% of a weighted average basis. Three, there is a momentum behind new LNG projects that could lead to several projects taking FIDs. Four, the priority shifts from offsetting CO_2 emissions to material carbon reduction; however, FIDs on capital-intensive measures (e.g., carbon capture and sequestration) are not likely. Five, global gas demand will remain resilient in the short term. Finally, natural gas is considered a transitional investment in the European Union taxonomy.

Petronas delivers Hiroshima Gas' first carbon neutral LNG cargo

Petronas' subsidiary, Malaysia LNG, delivered a carbon neutral LNG cargo to Japanese gas company Hiroshima Gas Co. According to CME Group, carbon-neutral LNG involves offsetting the carbon emissions from the LNG supply chain through the purchase of carbon offsets.

Canadian Natural Resources to raise capital spending this year

Canadian Natural Resources announced it will increase capital spending this year. The company cited sustained oil recovery and gas prices from pandemic-driven historic lows. Canadian Natural Resources will increase spending to more than \$3.4 B. The company also announced it will increase production.

BCCK awarded contract for Marcellus-Utica gas facility

BCCK Holding Co. was awarded an EPC contract to improve overall project economics at a cryogenic gas facility in the Marcellus-Utica Basin in southeastern Ohio (U.S.). According to BCCK, the company has developed an effective modification that will improve on the existing cryogenic plant design and equipment, increasing propane recovery and associated revenue in the process. The project will utilize a skidded BCCK patent-pending design, which will be available to enhance propane recoveries at many of the existing 200-MMsft^{3}d facilities operating throughout the U.S.

Fincantieri begins construction on largest U.S. LNG barge

Fincantieri Bay Shipbuilding has started construction on the largest LNG bunkering barge ever built in the U.S. The 416-ft, 12,000-m^3 vessel is scheduled to be completed in late 2023.

Once completed, the vessel will be under contract with Crowley, the largest independent operator of tank vessels in the U.S. Crowley will operate the vessel under a long-term charter with Shell. The vessel was designed by Crowley Engineering Services, the company's naval architecture and marine engineering solution group.

The ship will be used along the U.S. East Coast to expand current LNG network capacity and satisfy demand for cleaner energy sources for ships. LNG fuel is one way to help decarbonize the shipping industry.

McDermott wins major construction contract for Scarborough project

In January, Woodside awarded McDermott an engineering, procurement, construction, installation and commissioning (EPCIC) contract for a floating production unit for the \$12-B Scarborough LNG project. The floating production unit will process natural gas from offshore fields. It includes gas separation, dehydration, compression, mono-ethylene glycol regeneration and produced-water handling.

According to McDermott, the integrated scope also includes the design, fabrication, integration, transportation and installation of the hull and topsides. The 30,000-t topside will be fabricated by McDermott's JV fabrication yard, Qingdao McDermott Wuchuan, in China. The project scope includes a battery energy storage system to reduce emissions on the topsides and support Woodside's net emissions reduction targets.

Designed for a production capacity of up to 1.8 Bsft^{3}d, the topside will be connected to the semi-submersible hull and pre-commissioned prior to transportation and installation in a water depth of 3,100 ft (950 m), nearly 250 mi offshore Western Australia.

According to Woodside's website, the Scarborough project will extract natural gas from the Carnarvon Basin offshore Western Australia and deliver it via a 430-km pipeline to a proposed second LNG train (Pluto Train 2) at the Pluto LNG facility.

TotalEnergies signs agreements for the development of low-carbon natural gas projects in Oman

TotalEnergies has signed several agreements with the Sultanate of Oman for sustainable development of the country's natural gas resources. According to TechnipEnergies, the agreements include:

- The development of Marsa LNG, an integrated company between TechnipEnergies (80%) and Oman National Oil Co., OQ (20%). The JV includes the possible development of a low-carbon LNG plant in Sohar to produce LNG bunker fuel.
- The development of Block 10 natural gas fields in the western side of the country. Marsa LNG will hold a 33.19% interest in Block 10, with the project partners OQ and Shell Integrated Gas Oman B.V. holding the rest.
- A gas sales agreement under which Marsa LNG will sell natural gas from Block 10 to the government of Oman for a duration of 18 yr or until the startup of the Marsa LNG plant.

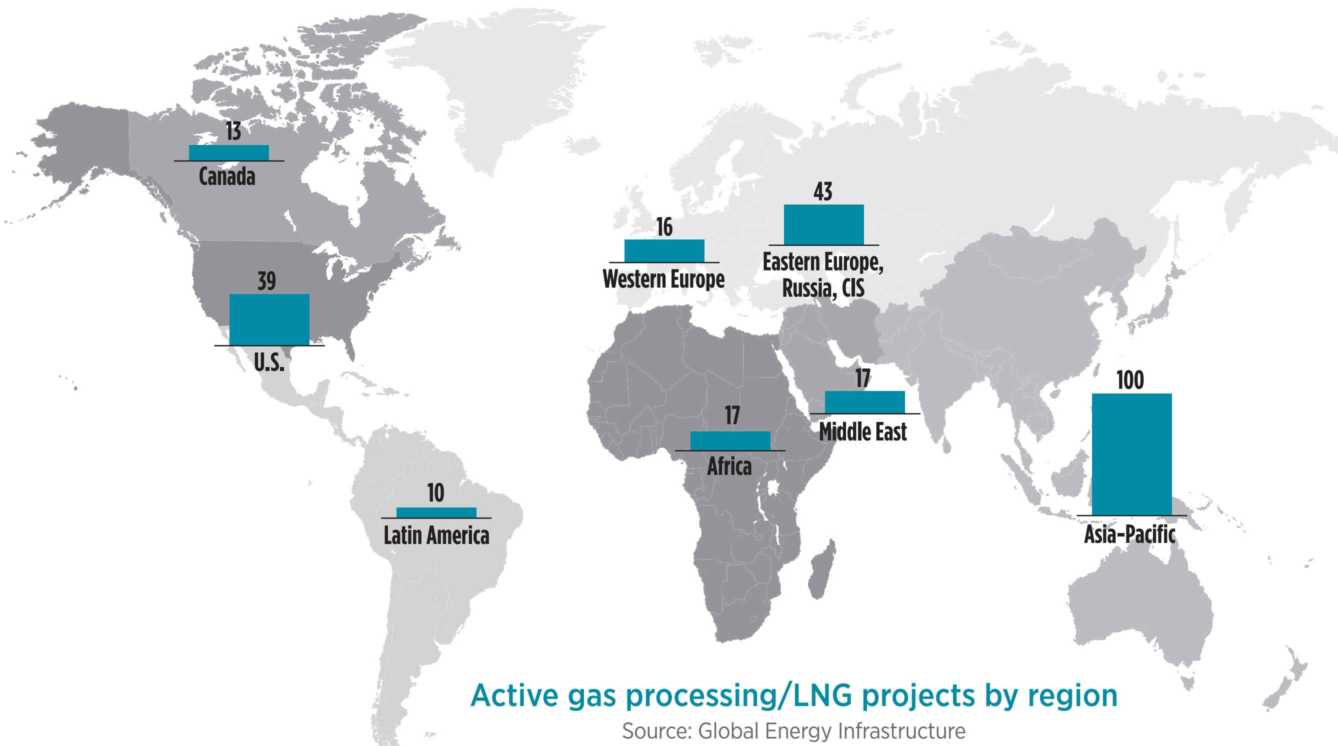
Brazil imports record LNG in 2021

According to Petrobras, Brazil imported 23 MMm^{3}d of LNG in 2021. This is a 200% increase in total LNG imports vs. 2020. Severe droughts in the country led to a deficit in power generation by hydroelectric power. To satisfy energy demand, Brazil was forced to import natural gas from countries like Trinidad and Tobago, the U.S. and Qatar. These supplies help supplement deficits for power generation.

Borouge to help power Egyptian Decent Life project

As part of Egypt's \$38-B Decent Life project, Borouge is collaborating with Egypt Gas to provide energy and infrastructure solutions to power thousands of households in Egypt. Borouge will provide gas distribution pipes to help Egypt build a safe and reliable gas distribution network.

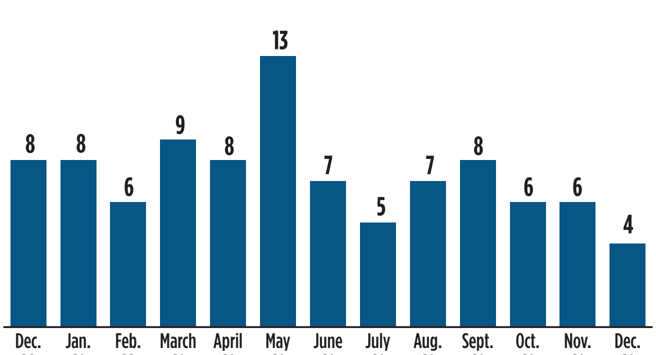
At more than 250 active projects, the gas processing/LNG sector represents nearly 25% market share in active projects globally. Most of the sector's active projects are in the Asia-Pacific region, followed by Eastern Europe, Russia and the Commonwealth of Independent States (CIS), and the U.S. Led by capital investments in China, the Asia-Pacific region has continued to dominate in total active project market share. China, along with several other Asian nations (e.g., India), are investing heavily in new LNG import infrastructure to satisfy increasing demand for natural gas. This is being led by new government initiatives to curb carbon emissions for power generation. **GP**



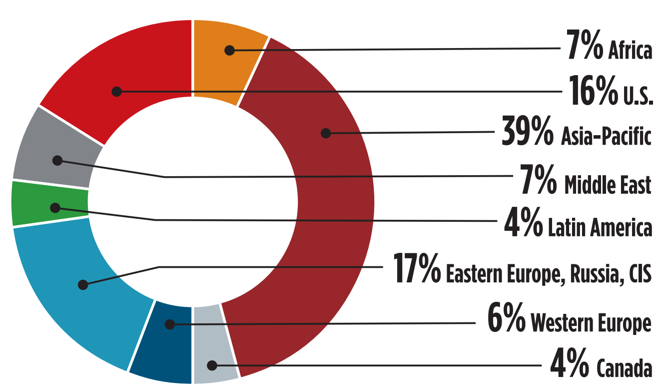
Active gas processing/LNG projects by region

Source: Global Energy Infrastructure

New gas processing/LNG project announcements, December 2020–December 2021



Active gas processing/LNG market share by region



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Russia ready to control 20% of global LNG market by 2030

E. GERDEN, Contributing Writer

Russia plans to increase its annual LNG production to 140 MMtpy by 2035. This increase in production—part of Russia's plans to diversify natural gas supplies—will enable the country to increase its global LNG production market share to 20%. The implementation of these plans will be part of the existing long-term program for the development of LNG production in Russia.

The program also includes additional production of LNG equipment and machinery within Russia, the expansion of a resource base for LNG projects and the provision of economic incentives for exporters. The program eases the procedure of issuance of licenses for the development of gas fields, which serve as the resource base for LNG facilities in the country.

The primary resource base for LNG production in Russia will be the Yamal and Gydan fields—which were discovered by Rosneft—and the Tambey group of fields discovered by Gazprom.

In 2020, Russia produced 30.5 MMt of LNG, which is an increase of 3.5% from the previous year. Out of the LNG produced, 22 MMt were exported. According to Russia's LNG export plan, both production and LNG exports should significantly increase through the rest of the decade. Forecasts show that Russia's LNG exports could reach $150 B by 2030. The increase in LNG exports will be due to new LNG facilities under development primarily in the country's northern region. The construction of these plants are being led by Novatek and Gazprom.

Gazprom has announced plans to build at least two medium-tonnage LNG plants by 2025. These projects are the $2-B–$2.3-B, 1.5-MMtpy Vladivostok LNG terminal to be built in the Vladivostok region (Russia's Far East) and the $500-MM–$2-B, 500,000-tpy–1.5-MMtpy Chernomorsky LNG terminal on the Black Sea. The Chernomorsky LNG terminal will supply LNG to countries on the Black Sea and Mediterranean Sea, as well as to enterprises of the South and North Caucasian Federal Districts in Russia.

At present, Gazprom's LNG portfolio includes only one operating LNG plant—Sakhalin-2. The 9.6-MMtpy Sakhalin-2 facility receives feed gas from the Piltun-Astokhskoye and Lunskoye fields. The LNG plant plans to add a third LNG liquefaction train, which will increase Sakhalin-2's total production by 5.4 MMtpy.

Liquefaction technologies, equipment and logistics. Gazprom needs foreign gas liquefaction technologies for its LNG facilities since it does not have its own LNG technology. To combat this, the Russian government plans to invest more than $1.7 B on new technologies for the domestic LNG sector, as well as deepening localization of LNG equipment within the country. As part of these plans, Russia plans to establish a production sector to focus on producing at least 18 of the most critical types of LNG equipment.

In addition to new equipment, the country will invest in new logistics and transportation infrastructure that will be crucial for Russian LNG exports to major foreign demand markets. According to Russian Prime Minister Mikhail Mishustin, the existing state strategy involves the establishment of specialized logistics centers (hubs) in the Russian Arctic region. These hubs will focus on the transshipment, storage and trade of Russian LNG, as well as future deliveries to demand centers in Europe and Asia. This investment will help Russia compete against major LNG exporting nations such as Australia, the U.S. and Qatar.

Russia's primary focus is satisfying increasing demand for natural gas in the Asia-Pacific region. Russia plans to supply the region with LNG produced from its Arctic plants. One of the key players in this region is the privately-owned gas producer Novatek.

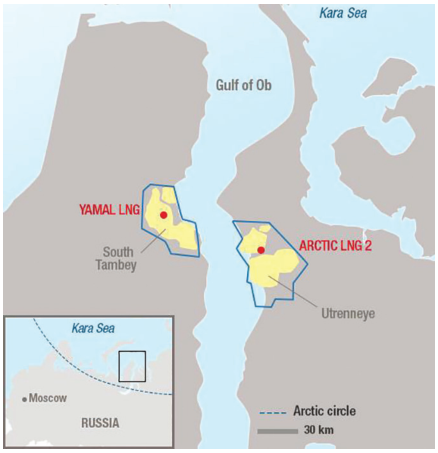

FIG. 1. The Arctic LNG 2 plant lies on the Gulf of Ob across from the Yamal LNG export terminal. Source: TotalEnergies.

Novatek is actively developing its Yamal cluster, which includes its flagship 20-MMtpy Yamal LNG plant. At the time of this publication, nearly 80% of Yamal LNG output has been supplied to Europe. However, to increase natural gas supplies to Asia, Novatek is developing a transshipment hub on the Kamchatka Peninsula. Scheduled to be completed in 2022–2023, the hub will enable Novatek to store and send LNG via ice breaking marine vessels—Russia's Zvezda shipyard plans to build up to 15 icebreaking tankers for Russian LNG deliveries—to demand centers in Asia. The company is also developing the nearly 20-MMtpy Arctic LNG 2 project. Located on the Gulf of Ob across from the Yamal LNG facility (**FIG. 1**), the three-train LNG plant is scheduled to begin operations in 2024–2025. **GP**

EUGENE GERDEN is an international contributing writer specializing in the global oil refining and gas industry. He has been published in a number of prominent industry publications.

The best ways to deal with heavy hydrocarbons, oxygen and helium in LNG plant feed gas

M. PATEL, Cheniere Energy, Corpus Christi, Texas

Natural gas is one of the world's most important sources of energy. Where natural gas pipelines are infeasible or do not exist, LNG is a way to move natural gas from producing regions to demand markets. Several recent LNG projects are based on pipeline gas (lean feed gas), which contains higher methane, with low natural gas liquid (C_2–C_5) and lower C_5+ heavy hydrocarbon content than typical conventional natural gas. It is very challenging to remove a small amount of impurities to ensure a stable and efficient operation.

Where does heavy hydrocarbon come from? Natural gas is available from conventional natural gas reservoirs and unconventional gas, such as shale gas, tight gas and coalbed methane. There are two primary sources for feed gas to an LNG plant: dedicated natural gas reservoirs and pipeline gas from mixed sources.

Hydrocarbon streams produced at the wellhead are composed of gas, liquid hydrocarbons and, sometimes, free water. The liquids from the gas phase are separated by passing the well stream through an oil-gas or oil-gas-water separator.

For pipeline gas, the removal of C_2+ heavy hydrocarbon components from natural gas is required to avoid the unsafe formation of a liquid phase during transportation. Therefore, pipeline gas is usually lean. However, a small amount of aromatics—such as benzene, toluene and xylene (BTX)—and C_5+ often remain in the gas. The typical composition of natural gas is detailed in TABLE 1.

What are the possible effects of trace amounts of heavy hydrocarbons in the lean natural gas feed of an LNG plant? The presence of heavy hydrocarbons in natural gas can result in a freezing out of the gas at liquefaction temperature, causing a significant adverse performance impact. Even trace concentrations of heavy hydrocarbons and aromatics can cause precipitation of solids (freezing) and fouling of the main liquefaction heat exchangers. When the freeze-out of heavy hydrocarbons occurs inside the exchange cores, pressure drop increases across the cores. This leads to flow maldistribution in the cores, which results in thermal stresses and core leaks.

When the core pressure drop reaches a high limit, the LNG train has to be shut down to defrost the exchangers and remove the heavy hydrocarbon deposits. Defrosting results in production loss and increases process risks such as equipment failure.

Hydrocarbons can also create operational and performance problems in the amine unit or the reinjection system.

What can be done to mitigate the effects of heavy hydrocarbons? This trace of heavy components must be removed prior to liquefaction. A few conventional separation schemes are available that use a scrub column and a liquids extraction unit that is integrated with the liquefaction process. These conventional technologies are often not feasible due to the low amount of heavy components. In addition, front-end natural gas liquids recovery processes may be feasible; however, an increase in capital and operational expenditures is associated with these configurations for the removal of only a small content of aromatics and heavy hydrocarbons.

A proprietary adsorption technology[a] is one of the most feasible solutions for lean gas natural gas processing. The adsorption technology is specifically designed for the removal of aromatic hydrocarbons (BTX) and heavy hydrocarbons from lean gas feeds in LNG pre-treatments (FIG. 1). This advanced technology combines both unit functionalities into one system by utilizing a multi-material approach to achieve heavy hydrocarbon and water removal to the required cryogenic specifications. This approach provides a 30% increase in capacity for BTX components, allowing for the effective removal of BTX to less than 1 ppm without compromising unit efficiency.

Natural gas extracted from conventional natural gas and/or oil fields does not generally contain oxygen (O_2). However, it is common to find certain high O_2 concentrations in U.S. natural gas pipelines (greater than 10 ppmv) in some

TABLE 1. Typical composition of natural gas

	Rich-feed gas composition, mol%	Lean-feed gas composition, mol%
Methane	85–90	96–98
Ethane	3–10	1–3
Propane	1–5	0.1–0.4
i-Butane	0.5–2	0.005–0.01
n-Butane	0.5–2	0.005–0.01
i-Pentane	0.5–2	0.005–0.01
n-Pentane	0.5–2	0.005–0.01
C_6+	0.5–2	0.005–0.01

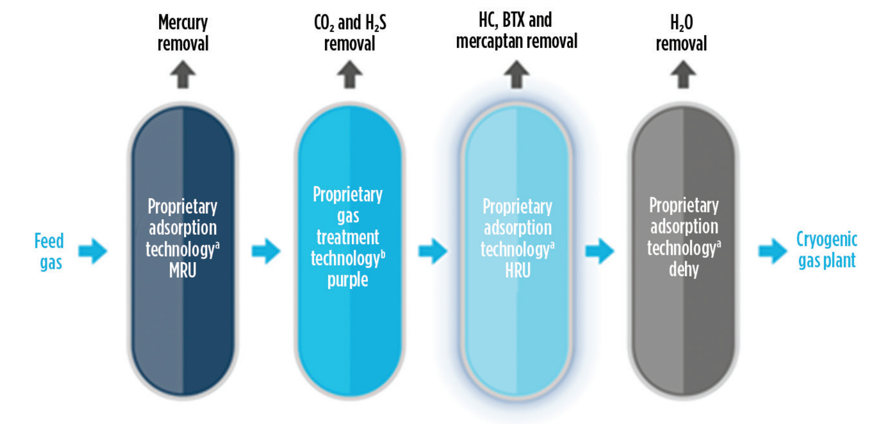

FIG. 1. The proprietary adsorption technology[a] is specifically designed for the removal of aromatic hydrocarbons (such as BTX) and heavy hydrocarbons from lean gas feeds in LNG pre-treatments.

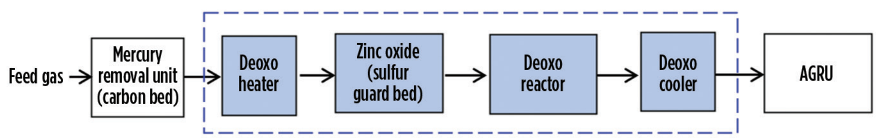

FIG. 2. A typical feed gas O_2 removal unit.

sampling analyses (where pipeline specifications are normally set at 10 ppmv). Usually, the presence of O_2 in natural gas sampling measurements is due to poor sampling procedures or to misread gas chromatograph results (it is not uncommon to confuse argon with O_2 when reading gas chromatograph results). It is also proven that some samples show much greater than 20 ppmv O_2 content, and most pre-treatment licensors can design systems to handle up to 20 ppmv of O_2. In some cases, the O_2 content may show values of 100 ppmv.

Helium is a very valuable gas in the industry. It is extracted from natural gas, which typically contains 1 vol%–8 vol% helium. While O_2 is a problem, helium might be a blessing. The following will analyze the challenges with O_2.

Where does that O_2 come from? Natural gas produced in conventional wells (either gas or oil) does not contain O_2, which can be found in gas lines fed by landfills, vacuum oil recovery systems, coal mine methane (CMM) or other low-pressure gas collection systems—all of which can contain several percent of O_2. Other sources of O_2 can be related to air ingress from field compressors in natural gas extraction systems (i.e., packing on low-pressure well compressors) or through ingress from upstream oil and/or gas production facilities (i.e., through tank vacuum breakers and inadequate control in vapor recovery units of storage tanks or O_2 rejection from helium recovery units) failure or lack of tank blanketing systems.

What can O_2 in natural gas feed do in an LNG plant? Even at low levels, the presence of O_2 in natural gas can result in the corrosion of pipes in LNG plants, which can cause safety hazards due to catastrophic gas leaks. O_2 can potentially impact the acid gas removal unit (AGRU) or the molecular sieve dehydration system. In the AGRU, O_2 can lead to a continuous formation of contaminants (bicine) by the degradation of the amine in methyl diethanolamine systems.

The reaction of O_2 with amines leads to the formation of different acids (i.e., formic, glycolic and oxalic), in which ions will form salts stable to heat. These salts increase the risk of foaming and fouling in the amine system, causing change in the amine strength and reducing the treating capacity of the amine system. Therefore, the AGRU will require more actions to maintain the amine quality.

In the dehydration unit, O_2 in the regeneration gas will result in the formation of water due to the oxidation of the light hydrocarbons during the regeneration heating stage. When present in excess of design limits, O_2 can cause problems in the dehydration unit because of the high temperature during the regeneration step. At the regeneration temperature, O_2 can produce an oxidation reaction that would generate water. If the natural gas also contains sulfur, the O_2 present in the regeneration gas stream will lead to the formation of sulfur dioxide (SO_2). Adding the presence of water will lead to hydrogen sulfide (H_2S) and sulfur. Sulfur can cause adsorbent deactivation and corrosion of materials in cooler parts of the regeneration section, as well as channeling and high pressure drop in the molecular sieves.

What can be done to mitigate the effects of O_2? The simple answer would be to remove O_2 from the feed gas (**FIG. 2**). In this sense, not many commercially proven O_2 removal systems exist in LNG plants. O_2 removal systems for LNG applications are based on catalytic reactions and solid scavengers (although there are not many references for the latter). A continuous adsorption-based process—using copper as the adsorber in at least two columns—could be used, but a reducing agent (hydrogen) would also be needed, which is very costly and not available in LNG plants. Therefore, we are left with systems based on catalytic reactions. For this option, the catalytic combustion of O_2 can be used to remove O_2 from natural gas. The combustion of O_2 in natural gas yields water and carbon dioxide (CO_2). Produced CO_2 will be removed in the downstream amine unit. Based on 50 ppmv of O_2 in the feed gas, there will be less than 50 ppmv of CO_2 in the feed to the AGRU, which will have minimal impact on AGRU operations.

These systems consist of a catalytic reactor with enough catalyst to remove the O_2 and produce CO_2 and water. The water could ultimately be separated in a separator or coalescer. However, it does not seem to be a feasible solution in LNG plants that might expect O_2 to be present in the feed gas.

Membrane technologies, like the ones used to remove CO_2 from natural gas, have been considered in some projects. However, there are no commercial membrane materials with proven results for the removal of O_2 from the natural gas.

Operational procedures in the LNG plant can be implemented to deal with O_2 contamination. These include the following:

- Consider the substitution of amine in the AGRU. The O_2 reacts with amines to form bicine and heat stable salts (HSS). Mitigation would occur by removing amine, using a portable reclaiming facility in the AGRU. The reclamation can be achieved with a mobile or permanent unit. When properly implemented, ion exchange technology has proven successful in removing HSS and bicine.
- In this process (FIG. 3), the lean amine solution is pumped into the ion exchange unit. The ion exchange resin removes the HSS, and the purified amine solution is returned to the amine circuit. After several minutes of operation, the system rinses the amine, using demineralized water from the resin, and a regeneration sequence takes place.
- To avoid possible operational problems in the dehydration unit, some dehydration unit licensors recommend staging the heating procedure to minimize the time in which the O_2 would be exposed to the high regeneration temperature. The idea is that, since most of the regeneration step occurs at mid-temperatures (around 340°F), the regeneration cycle is maintained at the same temperature. The final step is to heat the oxygen to high regeneration temperatures (445°F).

A combination of these two operational modes could handle up to 50 ppmv of O_2 content in the composition of the feed gas.

Should the O_2 content exceed 50 ppmv, the only feasible permanent solution is the installation of a closed-loop regeneration system in the dehydration unit. This solution falls within the realm of licensed designs. In this manner, there is no impact to the dehydration unit, and the regeneration operation would be under normal conditions (i.e., no high O_2 present). These types of installations would increase the regeneration needed for each dehydration bed; however, if the standby time is large enough, the overall dehydration-regeneration sequence would not be affected.

Helium recovery. Should the natural gas be rich in helium, the helium can be recovered in an integrated helium-nitrogen rejection unit in the LNG plant. However, in cases where the helium concentration is low, it is not economically feasible to extract it out of the natural gas stream.

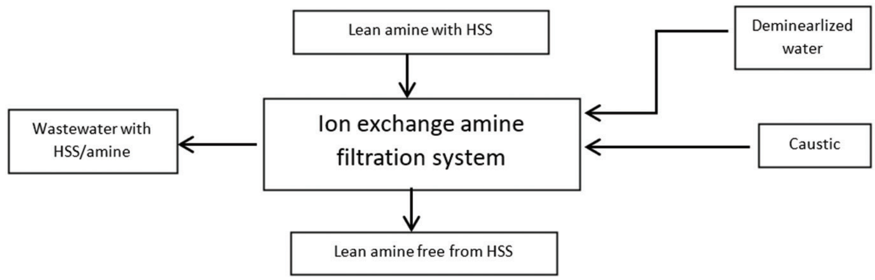

FIG. 3. Ion exchange filtration.

How can the extraction of low-concentration helium be economical in LNG plants? Boil-off gas (BOG) from LNG storage tanks is usually used as fuel gas for gas turbines in liquefaction plants (or for thermal oxidizers in plants that have electric motors as drivers). Besides methane, BOG also contains nitrogen, hydrogen, helium and some other heavier hydrocarbons. Depending on the case, a helium extraction unit can be installed in the BOG stream upstream of the BOG compressor; however, the purity requirements to export helium (> 95%) might require installing a hydrogen removal unit downstream of the helium extraction unit.

At present, there are several solutions to extract helium from natural gas. These solutions include membrane separation, pressure swing adsorption (PSA) and cryogenic processes. The latter seem to be the most economical solutions for this type of plant.

BOG components have different boiling temperatures; therefore, processes based on condensation and/or distillation seem to be adequate for this purpose. Generally, and from the perspective of producing helium with the required purity, high condensation pressures would be required to achieve such purity. In distillation processes, higher pressures lead to higher power consumption due to higher condensation duties.

A combination of both condensation and distillation is the most adequate solution to optimize the feed stream temperature and pressure of the cryogenic distillation column. The possible integration of a device to cool down the helium to condense the BOG to a lower temperature would provide favorable stripping through the condensation separation process at a relatively lower pressure. The low operating pressure would decrease the condenser duty and, thus, the power consumption of the system.

Takeaways. Even small amounts of heavy hydrocarbons in lean gas can be of critical concern for stable and efficient operations. While conventional technologies for lean natural gas processing are often infeasible due to the low amount of heavy components, the new adsorbent technology[a] is one of the most feasible solutions for lean natural gas processing.

The presence of O_2 in natural gas is problematic above a certain content; however, before considering any O_2 removal installation or change in operational procedures, it is imperative to confirm the existence of this gas within the natural gas stream.

While the extraction of helium also adds complexity to LNG process facilities, the payback period of such installations is very short. The overall profit of the LNG plant will be increased due to additional revenues generated from exporting extracted helium. Taking this into account, helium is a prize that can be well worth the extra effort and investment to obtain. **GP**

NOTE

[a] BASF's Durasorb Cryo-HRU technology
[b] BASF's OASE technology

MADANMOHAN PATEL is a Registered Professional Engineer in Texas and Louisiana, with more than 17 yr of chemical and process engineering experience in the LNG, petrochemical, chemical, oil and gas, and power industries. His experience is in supporting plant operational needs and in providing process expertise to troubleshoot and mitigate different plant issues. Mr. Patel earned an MS degree in chemical engineering from Texas A&M University.

Feed gas quality limits acid gas removal unit capacity

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Plant A is a gas processing and natural gas liquids (NGL) recovery facility designed to process 2.53 Bsft³d of sour associated gas received from four segregated gasoil separation plants (GOSPs). The separated gas is a mixture of solution gas (originally in crude) and cap gas (present in an upper dome over the crude and representing more than 60% of the separated associated gas from the GOSPs).

The sour gas is processed for NGL recovery, and some 1.5 Bsft³d of residue gas is recycled back to the GOSPs for reinjection. 260 MMbpd of C_2+ NGL is sent to Saudi Aramco's downstream facility through pipelines for further processing. Plant A consists of the following main process areas (FIG. 1 shows an overall process description):

- Inlet facility
- Acid gas removal units (AGRUs)
- Molecular sieve dehydration units
- NGL recovery units
- Acid gas compression
- Reside gas compression
- Injection gas compression.

Sour feed gas streams from the GOSPs combine in a common manifold and enter the Plant A inlet area, which consists of three parallel slug catchers with a capacity of 950 MMsft³d each. These slug catchers were designed as a three-phase separator to handle the slug during the pipeline scrapping operation and to separate the accumulated water and hydrocarbon liquids from the sour gas. According to the original design of the gas plant, no free liquid was anticipated during normal operation.

Sour gas from the slug catchers is joined again through a common manifold before it splits into two streams to feed the two identical AGRUs, which have a capacity of 1.253 Bsft³d each. The AGRU uses formulated methyl diethanolamine (MDEA) with piperzine (PZ) (50 wt% MDEA, 4 wt% PZ and 46 wt% water) to achieve a sweet gas specification of 4 ppmv hydrogen sulfide (H_2S) and 50 ppmv carbon dioxide (CO_2).

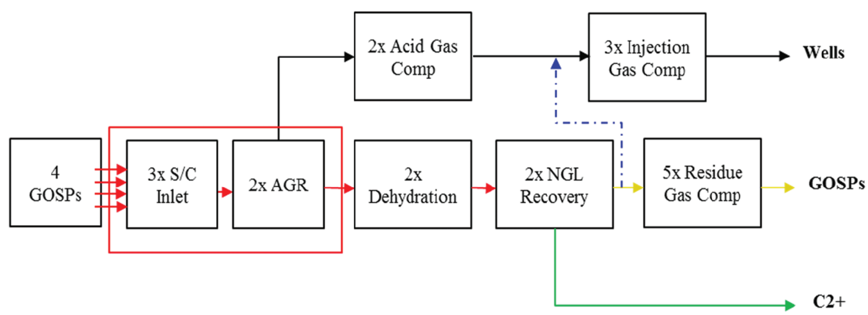

FIG. 1. Overall process description.

Sour gas is fed to the amine contactor where treated sweet gas is routed to the NGL recovery units for further processing. The rich amine is sent to the amine regenerator where acid gas is stripped off and sent to acid gas compressors for reinjection into the reservoir. The regenerated lean amine is recycled back and cooled by fin-fan coolers before it enters the top section of the amine contactor to remove acid gas from the feed sour gas.

Since its startup, Plant A experienced unstable feed gas quality that consequently impacted plant sustainability and resulted in capacity limitations and frequent foaming incidents. An engineering team performed extensive troubleshooting and investigation efforts to address these challenges, along with an action plan to mitigate the impact. The plant evaluation revealed that unit capacity limitation was mainly caused by increased acid gas content (H_2S and CO_2) in the feed sour gas by more than 50%. The presence of heavy hydrocarbons in the feed sour gas contributed to the foaming incidents in the acid gas absorber. Foaming resulted in excessive amine losses and off-specification product gas, as well as sudden reductions of feed gas flows. The issue was more severe during summer, when the unit experienced several incidents of foaming on a weekly basis. The impact resulted in more than a 50% reduction in plant processing capacity on a temporary basis.

Methodology and study basis. The comprehensive engineering efforts to evaluate and troubleshoot the capacity limitation of Plant A considered the following during the assessment:

- Review of the plant design, including the basis of design and original simulations.
- Review of the inlet area, AGRU and upstream GOSPs operation, which included main parameters changes, equipment failure and main streams conditions.
- Process data from the inlet area, AGRU and upstream GOSPs. This included historical process input and trends, operations records, routine/non-routine streams analyses, and visual inspection of main equipment in the field during operation.
- Collecting different gas and amine samples and performing simulation based on design and actual data.

FIG. 2. Hydrocarbon layer in an amine contactor.

FINDINGS AND ANALYSIS

Plant design and simulation. A thorough review of Plant A design revealed the following:

- The field has cap gas that represents more than 60% of the produced gas at the GOSPs. This cap gas is mixed with the crude inside the reservoir. During plant design, the GOSPs were simulated based on crude oil samples representing the north and south part of the field; however, only the cap gas composition for the south field was used. This resulted in lower acidity in the final simulation. Late samples revealed that the north field cap gas has a higher acid gas content; therefore, it has a major impact on produced gas acidity.
- The compositions and flowrates of the gas produced from the GOSPs are not identical. To some extent, more gas is produced from the north field (hence the higher acid gas content).
- The gas simulation of the GOSPs without the north field cap gas data resulted in low acid gas content (1.2 mol%–1.4 mol%) in the GOSPs' associated gas. However, when the north field (higher acid gas content) data was used, the acid gas content in the GOSPs' associated gas increased to 2.1 mol%.

Upstream GOSPs and inlet area. Based on the team assessment of upstream units operation (GOSPs, inlet area and feed gas filter-separators), the main findings included:

- **Normal operation at all GOSPs.** Several factors contributed to periods with unexpected gas compressors shutdowns; however, no direct impact on the downstream facility could be noticed. The number of shutdowns was significantly reduced later.
- **Steady operation of the triethylene glycol (TEG) dehydration unit.** The four GOSPs are comprised of nine upstream TEG units. Their operation was steady except for one of the units, where low performance was noticed—this was due to operational issues in the TEG regeneration section. The impact of this will be related to water carryover with the gas, which has no direct contribution to foaming. The subject TEG unit performance was significantly improved later. The TEG units were excluded as a direct cause of the foaming issue.
- **High operating temperature of the TEG units at the GOSPs.** The GOSPs area experiences high ambient temperature during summer: it is difficult to cool the associated gas and maximize heavy ends condensation. This results in intermittent TEG carryover and high hydrocarbon dewpoint of the produced gas. The original facility was designed in the 1990s without NGL recovery, so no refrigeration unit was built. NGL recovery was introduced in 2015, so it was infeasible to install refrigeration units at the GOSPs.
- **High acid gas and C_6+ content.** Many gas samples during summer and winter were collected from the outlet of the four GOSPs, the inlets to the Plant A slug catchers and the inlets to the two AGRUs. All samples confirmed high acid gas content of 1.8 mol%–2.1 mol% of acid gas (H_2S and CO_2) and slightly higher C_6+ levels of 0.7 mol%

The presence of hydrocarbons in AGRU equipment. The pure amine solution does not foam easily, but foaming can be initiated by many factors, such as hydrocarbon liquid, corrosion inhibitor and amine degradation products. Plant A experienced several foaming events at different intervals in the amine contactors in both AGR trains. The foaming resulted in excessive amine losses and off-specification product gas, in addition to sudden reductions of feed gas flows. Frequent liquid hydrocarbon accumulation occurred in the amine contactor at both AGR trains. At peak times, the Plant A operation team drained the amine contactor several times per day to skim the liquid hydrocarbon that was accumulating in the system.

During the summer of 2017, numerous foaming incidents took place, so the team investigated the amine contactor and flash drum operations. It was confirmed that there was significant liquid hydrocarbon carryover to the AGRU unit. This was proven through:

- Visual inspection of the equipment sight glass, which showed a significant layer of hydrocarbons (**FIG. 2**)
- Excessive skimming of both the amine contactor and flash drum, multiple times a day
- Amine samples collected that showed the presence of hydrocarbons in lean amine solution.

Amine solvent quality. Monitoring the amine solvent quality is crucial to assess the AGR unit performance. The Plant A lab and several third-party labs analyzed the lean amine solution quality for both AGR trains. **TABLES 1** and **2** show the analyses results:

- H_2S and CO_2 in the lean amine for both trains are below the recommended range of 0.0005 mol/mol–0.003 mol/mol, which indicates over-stripping of rich amine. The presence of H_2S at low levels (typically 300 ppmv) is important to provide an iron sulfide protection layer against corrosion.
- The samples showed that amine solvent concentration (MDEA + PZ) is within the recommended range of 50 wt%–57 wt%.
- All analyses results showed TEG content in the amine samples, the impact of which is discussed here.

TEG contamination in the amine solvent loop has a negative impact on the gas treat unit capacity, as it can replace the active amine. The GOSPs TEG systems are designed to dehydrate the sour

TABLE 1. GT-1 amine analysis

	Sample 1, August 2016	Sample 2, December 2016	Sample 3, February 2017	Sample 4, April 2017	Sample 5, August 2017	Sample 6, September 2017	Sample 7, November 2017
MDEA, wt%	55.3	49.7	53.15	51.2	51.2	50.4	46.7
PZ, wt%	4.4	3.4	3.27	2.7	2.7	6	7.2
TEG, wt%	1.08	2.55	3.54	3.9	3.9	2.86	3.9
Lean loading, mol/mol	0.0005	0.0013	0.0009	0.0008	-	0.0025	0.0025

TABLE 2. GT-2 amine analysis

	Sample 1, August 2016	Sample 2, December 2016	Sample 3, February 2017	Sample 4, April 2017	Sample 5, August 2017	Sample 6, September 2017	Sample 7, November 2017
MDEA, wt%	40.4	42.7	49.9	46.8	46.8	47.4	48.1
PZ, wt%	4.6	3.9	5	4.6	4.6	6	6.7
TEG, wt%	-	1.8	2.29	2.8	2.8	2.9	3.9
Lean loading, mol/mol	0.003	0.0019	0.001	0.0018	-	0.0034	0.0017

gas and remove the water content below 7 lb/MMsft^3 to protect the pipelines from water condensation. The TEG contactor is operating at 140°F, where the expected TEG vaporization losses are relatively high.

High acid gas contents in feed gas. The authors' company's lab and a third-party vendor lab conducted extensive work to perform complete analysis of the sour gas from the four GOSPs plant outlet and the Plant A inlet. The measurements were conducted between November 2016 and October 2017. In addition, a plant simulation was performed using crude and gas samples representing the whole field (north and south sections) to verify and compare with actual measurements. The simulation results were validated by the field sample analysis. The following were practiced during field sampling:

- Samples were collected from four GOSPs outlets and the two AGRU inlet headers
- Two samples per day representing day and night conditions were collected for each location
- Samples were collected for at least one week. More than 100 samples were collected in total, representing summer/winter and day/night compositions of the sour gas.

Based on the samples analyses and actual simulation of GOSPs operation using the updated compositions:

- The measurements and simulations confirmed a significant increase of acid gas content in the feed gas stream (FIG. 3) from the design of 1.21 mol% to 1.8 mol%–2.1 mol%.

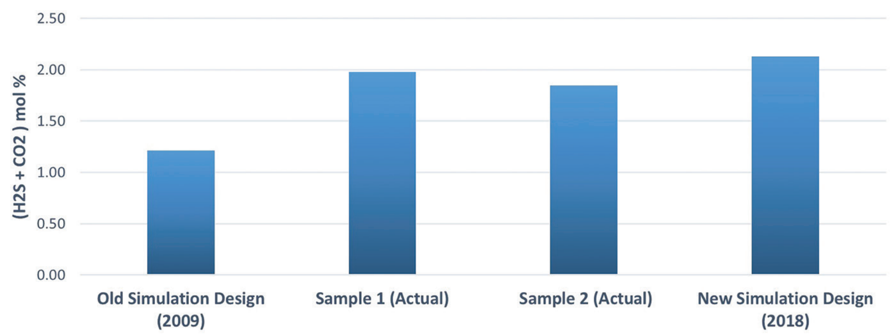

FIG. 3. Acid gas content in feed gas.

- The C_6+ hydrocarbon increased from 0.68 mol%–0.71 mol%.
- A significant BTEX (benzene, toluene, ethylbenzene and xylene) of more than 1,000 ppm was measured. The BTEX measurement was not a typical analysis during the design phase, therefore it was not considered. BTEX has a significant impact on promoting foaming in the amine contactor. Although the unit is operating at a relatively high temperature (140°F–145°F), which reduces BTEX absorption, the solvent-rich loading is relatively low (0.2 mol/mol–0.25 mol/mol), which promotes BTEX absorption.

Since the AGRU was designed for 1.21% mole acid gas (equivalent to 37 MMsft^{3}d acid gas volume), the new acid gas content of 1.8%–2.1% limits the plant processing capacity to less than 1.8 Bsft^{3}d, which is a reduction of more than 600 MMsft^3 of processing capacity.

The main conclusions from the study are:

- The plant feed gas composition that was used to design the AGRU did not represent the actual composition of the feed gas to the plant—mainly the acid gas content and heavy C_6+ components, which had significant impact on AGRU capacity. The actual sour feed gas acid content is more than 50% compared to the original design.
- Heavy hydrocarbons frequently collected inside the AGRU amine contactor. These hydrocarbons were the main cause for the frequent foaming in the unit.
- The topography of the plant area resulted in unstable feed conditions (temperature and liquid carryover) to the plant. Diurnal impact was evident were the feed gas temperature fluctuated significantly between day and night, especially from nearby GOSPs (2 and 3)
- The oil field has unique characteristics: the south part cap gas contains less acid gas

(~ 1 mol%), while the north section of the field has more acid gas in the cap gas (~ 2.3 mol%). This resulted in a major discrepancy between the actual feed gas composition and the original composition that was used during the design.

- The inlet gas temperature was too high and resulted in lower amine capacity and, consequently, limited the AGRU from accommodating higher feed gas.
- The current piperazine concentration of 4 wt% contributed to the limited amine capacity to process additional acid gas.
- The upstream equipment was inefficient in removing entrained liquids. The inlet slug catchers did not have internals to reduce liquid carryover, and the inlet feed gas filter separator size was inadequate to handle the design flowrate at actual conditions.

RECOMMENDATIONS

Install a dewpoint control unit (DPCU) upstream of the AGRU. The DPCU will eliminate the impact of fluctuating sour feed gas conditions (liquid carryover, temperature and composition). It will ensure stable and lower feed temperature to the AGRU, remove heavy hydrocarbons, significantly improving the performance of the AGRU and enable the unit to process more sour gas. The design of the DPCU considers multiple raw crude and gas cap samples representing the field.

The raw data is used to develop a new simulation of the plant based on the latest crude oil and cap gas composition, representing the north and south part of the field to design the new DPCU and debottleneck the current AGRU capacity limitation.

Although this option will imply capital investment, it is the ultimate solution to resolve the frequent foaming and limited AGRU capacity. It is critical to consider such an option at the early stage of plant design, bearing in mind all factors that affect the decision (i.e., actual gas hydrocarbon dewpoint temperature, content of C_6+, ambient conditions, length and size of feed gas pipelines).

The new simulation results and actual gas composition measurements will be used as the basis for any new plant modification and DPCU design (FIG. 4).

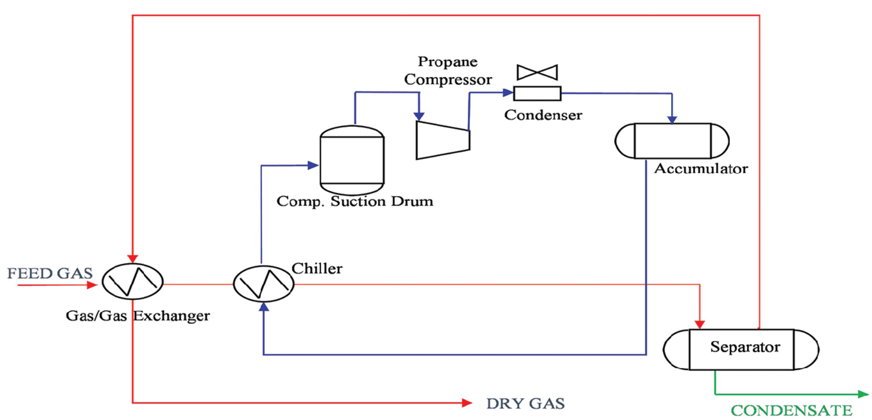

FIG. 4. Schematic of the proposed dewpoint control unit.

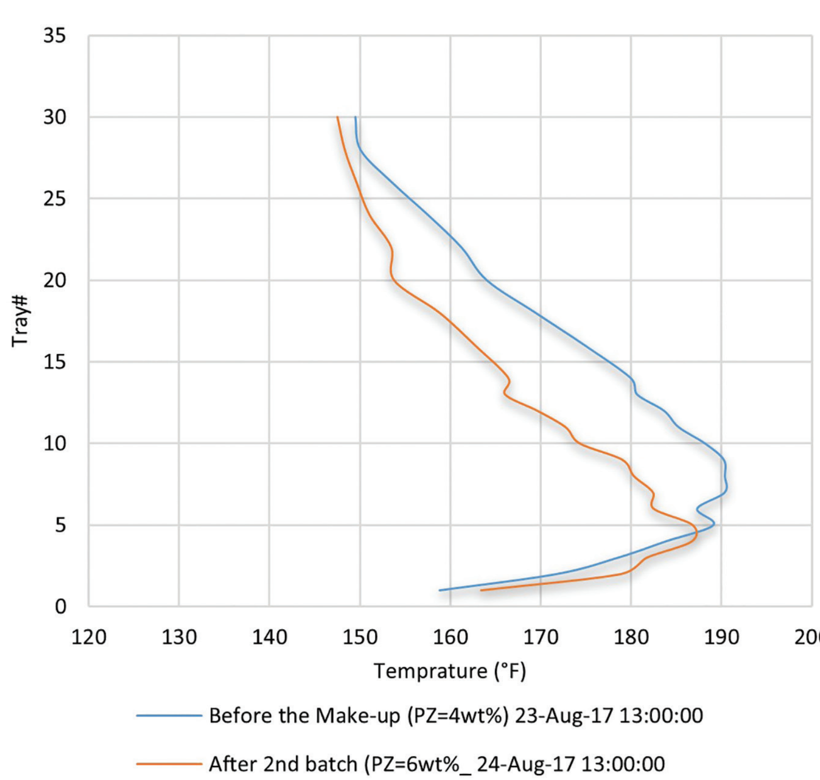

FIG. 5. Amine contactor temperature profile for GT#1.

Increase piperazine concentration. The increase of piperazine concentration will result in higher capacity (pickup ratio) and a lower contactor temperature in active trays, which will further increase unit processing capacity. In this case, the lower design piperazine concentration provided the flexibility to increase the concentration to support higher amine loading.

This proposal was simulated using process simulation software[a], and the new temperature profile showed that acid gas absorption is enhanced, which will allow the amine contactor (FIG. 5) to accommodate more acid gas content (a 4°F reduction in bulge temperature was realized).

Modify the slug catcher internal. The three slug catchers are the first protection layer to remove the heavy hydrocarbon from the associated gas fed via pipelines. The existing slug catchers were provided with conventional demister mesh pads, which proved to be inefficient and resulted in a significant amount of liquid hydrocarbon and water carryover with the gas stream to the amine unit. The situation was aggravated under the several operating scenarios, such as occasional slug flow

regimes (scrapping, startup/ shutdown), solids entailment in the feed gas and defective level controls. Therefore, it was proposed to modify the slug catcher internal to improve gas distribution, improve solid removal, enhance residence time, improve gas-liquid separation and minimize carryover to less than 0.1 gal/MMsft[3].

Install a gas-liquid coalescer. The feed gas filter-coalescer is an essential piece of equipment in the AGRU, as it protects the amine unit from any liquid carryover. Typically, the system should be designed with the highest removal rate (99.9% solid removal and maximum of 0.1 ppmv liquid in the gas outlet). Since the existing filter-separator is under-sized, a new filter-coalescer was recommended. An alternative is to upgrade the existing one through the use of different coalescing elements. The new conditions after installing the dew-point control will determine the need to replace or upgrade the existing system.

Use alternative anti-foam agent. The current anti-foam that is used at Plant A is silicone-based. During foaming upsets, batch injection (concentrated anti-foam) is applied. The excessive use of this type has a potential to create operational issues, such as accumulation in the system, which could reduce amine capacity and block equipment like the lean rich exchanger.

It was recommended to consider other types with less potential impact, such as glycol-based. The appropriate type should go through laboratory testing to ensure effectiveness, compatibility and minimum long-term effect.

A new type of anti-foam was implemented and based on the continuous monitoring of the main foaming related parameters of the AGRU, the units' foaming tendency was significantly reduced and the number of foaming incidents decreased. **GP**

NOTES

[a] ProMax

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Innovative technology for re-liquefying evaporated LNG to meet net-zero emissions targets

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In the current era, environmental challenges do not exist in isolation. The protection of the environment is one of the greatest challenges facing the world.

Many industries are swiftly moving toward the green environment, and many companies are adopting more stringent environmental regulations that require the provision for a vapor control system to reduce the emissions of volatile organic compounds (VOCs). These investments are to ensure improvements in air quality, resulting in a healthier environment and improving the health and well-being of its inhabitants.

Vapor recovery for LNG. In countries like Australia, vapor recovery has become mandatory in major urban areas. One technology that has evolved is re-liquefaction of evaporated LNG—boil-off gas (BOG)—on LNG carriers. The BOG released from LNG storage vessels on carriers can be used to power the steam turbines to run the ships, all the while reducing greenhouse gas (GHG) emissions. Advancements in turbo-expander technologies have led to the development of new re-liquefaction systems.

LNG usually comprises methane, ethane, propane, butane and nitrogen. LNG is sold on an energy basis—therefore, the composition of the product is of great importance. Nitrogen is normally removed from the natural gas stream through a nitrogen rejection unit to increase LNG's energy value. The energy value of LNG is measured based on the higher heating value (HHV), which is the amount of heat obtained by the complete combustion of a unit quantity of a material.

LNG is stored at approximately –163°C in tanks close to atmospheric pressure. The pressure in the tanks is normally less than 7 psig. LNG cryogenic storage tanks are equipped with extremely efficient insulation; however, heat is still transmitted due to the surrounding atmosphere. For the LNG to stay liquefied, it must stay at or near constant temperature, which can be achieved if the tank is held at a constant pressure.

To remain at a constant pressure, the LNG vapor boil-off is allowed to leave the tank, which will keep the pressure constant. BOG from the cryogenic tank is of high value; therefore, it is compressed and used as a fuel gas (not flared).

BOG may also be produced when LNG is pumped into LNG carriers. Due to the temperature of the empty tanks, BOG forms quickly. The rate that the BOG is produced begins to slow down quite substantially as the LNG and the tank temperatures reach equilibrium. BOG can be converted to a fuel source, or it can be re-liquefied in the stream and exported as LNG. Re-liquefaction systems can be optimized based on the following:

- Arrangements of the turbo expander
- The pressure of the BOG liquefaction system
- The refrigerant selection.

With the optimization of the turbo-expander arrangements (the refrigerant selection and the BOG feed pressure), energy savings of approximately 30% can be achieved.

Nitrogen liquefaction systems. There is a constraint on refrigerant at a pressure of 2 bara due to the necessity of latent heat at –155°C. Nitrogen is the only refrigerant that can efficiently liquefy the BOG at 2 bara.

FIG. 1 is a generic process flow diagram required for the liquefaction of BOG at 2 bara. The flow through the Joule-Thomson (JT) valve provides the latent heat required to liquefy the process fluid. The purpose of the JT valve is to provide liquid nitrogen expansion, which provides the latent heat required at the cold end of the system. The discharge pressure must be controlled to have an efficient system. When nitrogen is used as a refrigerant, the discharge pressure from the JT valve is acceptable. Because of the boiling temperatures of the components, a methane refrigerant has a discharge pressure near atmospheric pressure.

The LNG is cooled using liquid nitrogen, which is cooled by the nitrogen in the cold and warm turbo expanders. **FIG. 1** shows the use of a JT valve for latent heat cooling and a turbo expander in re-compressor mode to provide sensible heat for warm-end cooling. The following are the main points of this study (**FIGS. 2** and **3**):

- Sufficient heat is between ambient temperature and –155°C.
- Nitrogen refrigerant can provide latent heat at –155°C at realistic operating conditions (methane at 160 kPaA).
- The nitrogen and carbon system is more efficient.

With a combined refrigerant case, a turbo-expander loop will cool, using a nitrogen feed, and a second turbo-expander

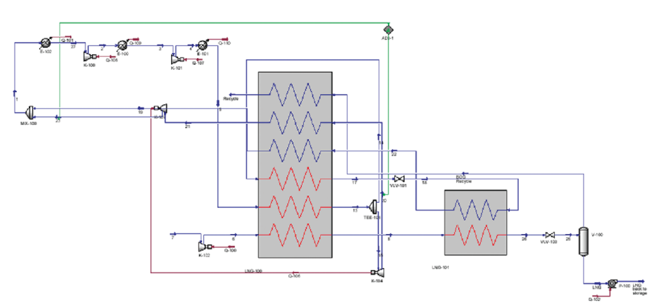

FIG. 1. Process flow diagram required for BOG liquefaction at 2 bara.

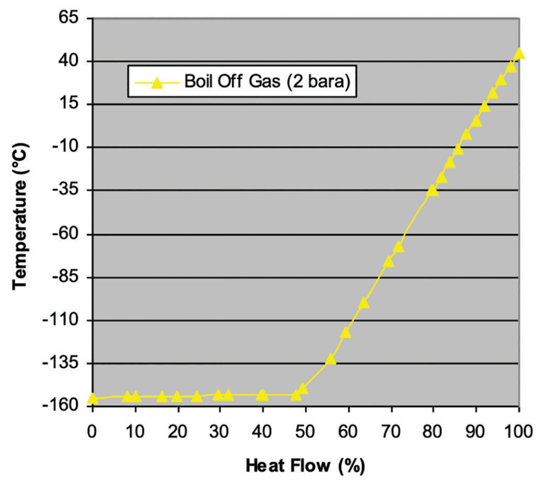

FIG. 2. LNG heating curve.

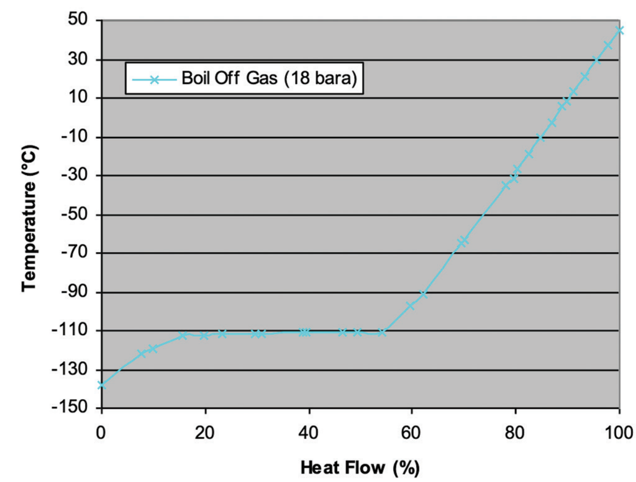

FIG. 4. LNG heating curve for combined refrigerant case.

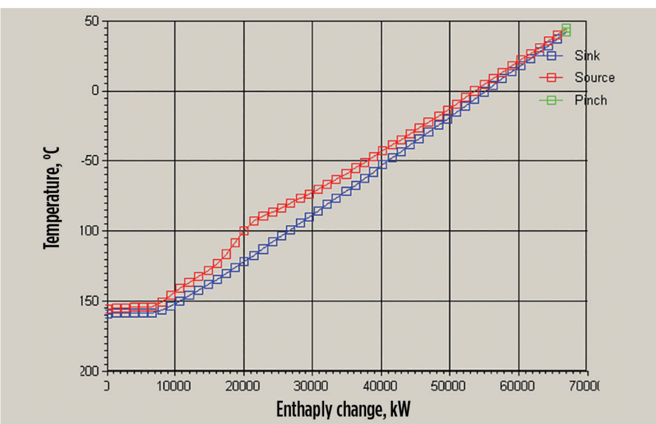

FIG. 3. Enthalpy change-temperature.

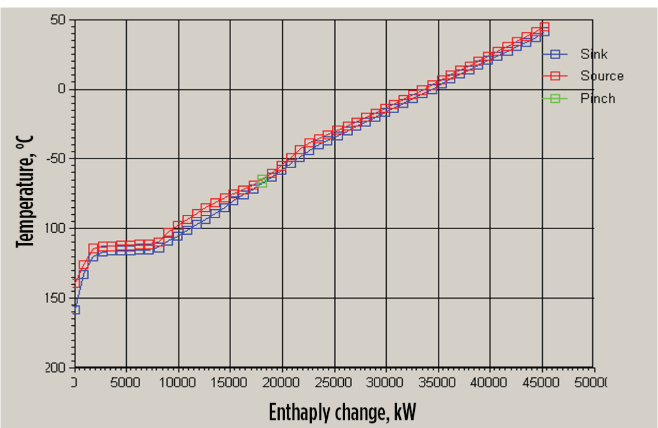

FIG. 5. Enthalpy change-temperature for combined refrigerant case.

loop will cool the warm end with a methane feed. The intermediary section—which requires latent heat—will be cooled with the use of liquid methane. The methane system at the cold end will use latent heat to provide the required cooling. The limitation is that methane will begin to condense in the temperature range between –110°C and –143°C (**FIGS. 4** and **5**). The details of this case include:

- The requirement of latent heat is –110°C.
- Nitrogen refrigerant can provide sensible cooling at –155°C.
- Latent heat is provided with methane at –110°C within realistic operating conditions.

The turbo-expander system is only able to follow a straight path; therefore, there will be areas of inefficiency at the significant points of curvature. Looking at systems at 18 bara and 2 bara, there is no real curvature. The changes in the curves are sharp, and this can be modeled using a turbo-expander cycle. At 50 bara, a mixed refrigerant could be used to follow the curves more efficiently (**FIG. 6**). The varying composition of methane, ethane and nitrogen means that the evaporation of the refrigerant will occur over a larger range of temperatures. The accurate selection of the mixed-refrigerant composition is critical to follow the cold end of the LNG heating curve. The problem with this system is that methane will still condense at low-end temperatures. The result will be a larger gap of inefficiency, which is significant. The process flow diagram and enthalpy change are shown in **FIGS. 7** and **8.**

Mixed-refrigerant liquefaction systems. The LNG heating curves play an important role in the selection of the refrigerant. Mixed refrigerant would be unable to achieve cooling along the flat sections of the heating curves. When the BOG is at 50 bara, there is more curvature in the heating curve. This means that the process fluid is condensing through a wider range. For this reason, a mixed refrigerant is suitable, since the components will also condense at different temperatures. A mix of the three components (methane, ethane and nitrogen) will result in a curve that will justifiably provide cooling at the cold end of the system.

The warm end of the system can be cooled with the use of heat. Therefore, the use of turbo expanders is appropriate. The turbo-expander cycle can use nitrogen, methane or mixed refrigerant as a refrigerant. As proved in the previous models, methane is a more effective refrigerant due to its molar enthalpy. In addition, ethane should be a better refrigerant than methane; however, its range of operation is limited.

The mixed refrigerant will be more efficient at the cold end due to the latent heat of vaporization that it possesses. If mixed refrigerant is only being used at the cold end, heavier hydrocarbons (such as propane and pentane) should not be used.

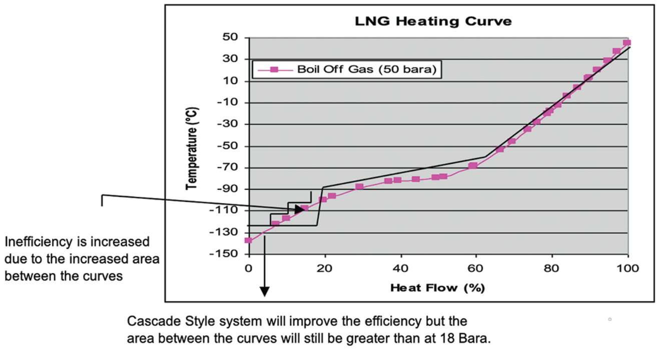

FIG. 6. LNG heating curve of turbo-expander system.

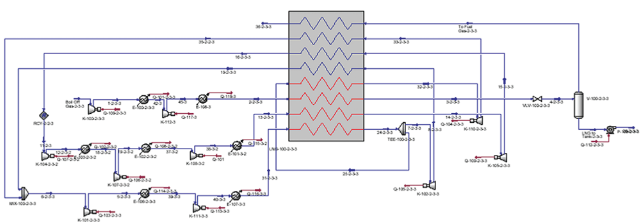

FIG. 7. Process flow diagram of turbo-expander system case.

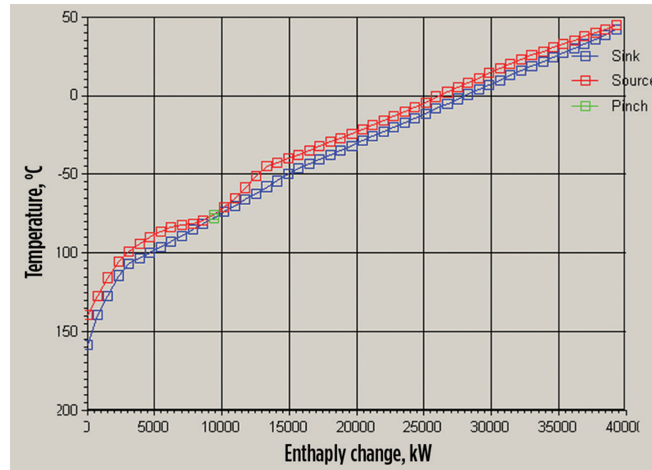

FIG. 8. Enthalpy change-temperature of turbo-expander system case.

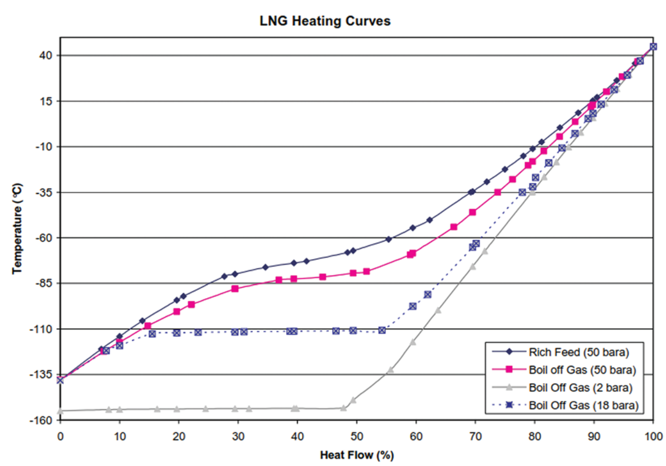

FIG. 9. BOG will liquefy more efficiently as a higher feed pressure.

Warm-end cooling can be achieved by using heat provided by expansion. Mixed refrigerant is more efficient than nitrogen at the warm end, as it possesses a higher molar heat capacity. Therefore, mixed refrigerant is normally used in the warmest turbo expander to provide the greatest efficiency within the operating limitations of the expanders.

With the optimization of the turboexpander arrangements, the refrigerant selection and the BOG feed pressure, energy savings of approximately 27% can be achieved.

The system's efficiency improves with increasing the BOG feed pressure. When the BOG feed pressure is increased, the heating curve becomes more curved. Condensation does not take place over a limited area, but rather across a larger area.

The BOG will liquefy more efficiently at a higher feed pressure. This is clearly presented by the respective LNG heating curves in **FIG. 9.** As the feed pressure of the BOG is increased, the curvature in the LNG heating curve begins to increase. The heating curves also show the required cooling.

This means that methane is more efficient than nitrogen due to the molar heat capacity it possesses. This was a significant advantage, and methane was used for all warm-end refrigeration. The challenge with methane was its ability to provide sensible heat at cold temperatures below –110°C. Due to methane having a warmer boiling point than nitrogen, it was unable to effectively provide cooling below these temperatures, using a turbo-expander cycle. At 18 bara, the evaporation of methane to provide cooling was still more effective than the combined nitrogen and methane system. At 50 bara, methane was inefficient due to its inability to follow the curve at the cold end. In this case, it was advantageous to have combined nitrogen and methane systems.

Further advantages included using ethane as a refrigerant, which has an even higher molar heat capacity than methane. The use of a mixed refrigerant with ethane then became quite attractive. Ethane has a higher boiling point and begins to condense at warmer temperatures. This restricts its use in turbo expanders, where the discharge stream must have less than 10% liquid flow. In the most efficient modeled system, the mixed refrigerant only passes through the warmest turbo expander to produce no liquid on the outlet of the expander. This still provides a significant energy savings of approximately 3% vs. methane being passed through the expander.

Therefore, methane was a more effective refrigerant than nitrogen since methane has a higher molar heat capacity. This makes the methane system more efficient than the nitrogen system. The same was found with the mixed refrigerant, which condensed at a higher temperature due to the ethane content. The mixed refrigerant has an even higher molar heat capacity than the methane due to the presence of ethane. This justified why the mixed refrigerant system—with mixed refrigerant flowing through the warmest expander—was the most efficient system.

A large amount of energy could be efficiently recovered, and maximum efficiency of the whole process system could be achieved by applying the best engineering approach. Simultaneously, it would also contribute to reducing the exposure of GHG and to improving the air quality, resulting in advancements toward achieving net-zero emissions targets. **GP**

Optimum air exchangers fans control selection in distillation columns

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Air exchangers are commonly used in distillation columns to remove heat generated by the columns' reboilers, and use the lower temperature product as reflux. The target temperature in the air exchanger outlets can be controlled by several methods, such as louvers, pitch controls (on the fan blades) or variable-speed motors.

Failure to control the target temperature from the air exchangers can lead to energy losses, additional heat requirements and lost product specifications. Most air exchangers consist of several parallel units, covering a large area that makes it difficult to control the target temperature by solely depending on using air exchanger fans on-off (as a step-control).

Various target temperature controls for air exchangers can be made through the following applications:

- **Louvers:** A set of angled slats fixed or hung at regular intervals to allow air or light to pass through.
- **Variable-pitch mechanism:** Axial fans usually have a variable pitch—with blades that change pitch in operation—so that the amount of airflow is furnished to meet the requirements for a particular heat exchanger.
- **Variable-speed motor:** A variable frequency drive or similar technology that is installed to control motor speed and torque.
- **Fin fan step-control on-off:** The sequential shutdown or startup of fan motors to stop or allow airflow through exchangers.

Louvers tend to be large, awkward and easily damaged. The variable-pitch mechanism has been used for control but seems to feature deadband or hysteresis. The combination of air exchanger fan step-control (on-off) and variable-speed motor was determined to be a better choice for control.

In the following sections, a demonstration is detailed to select the optimum number of variable-speed motors for distillation columns in an NGL fractionation process (depropanizer column) while configuring the rest of the air exchangers fans to be switched on-off, based on their temperature control range limits.

NGL FRACTIONATION DEPROPANIZER DISTILLATION COLUMN

The depropanizer column is a distillation column used to recover the propane product from the natural gas liquid (NGL) feed stream to the column. FIG. 1 shows a typical layout of the column and associated equipment.

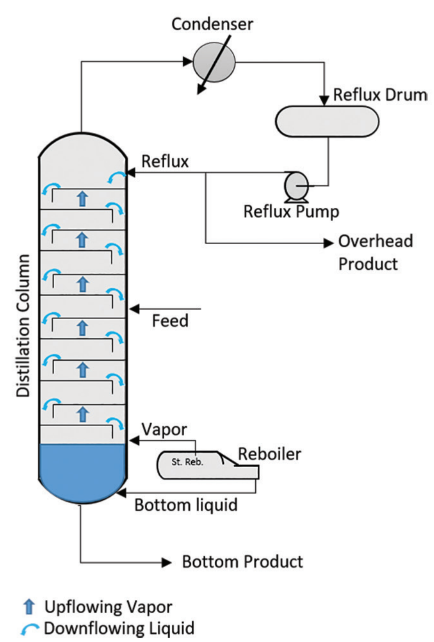


FIG. 1. A typical layout of a depropanizer column and associated equipment.

The selected depropanizer column contains a high number of air exchanger fans. The objective is to determine the optimum number of fans to be equipped with a variable-speed motor, and the rest will each be assigned a step-control (on-off).

To reach this objective, it is essential to define the minimum acceptable temperature (MAT) with the available reboilers and air exchanger duties. These minimum temperatures can be used while varying the ambient weather temperature to define the number of required fans.

Operation stability and MAT in reflux. In this section, the air exchanger minimum reachable temperature is determined by the duties available in the top and bottom sections of the column, by an air exchanger in the top section of the column and reboilers in the bottom section.

The bottom temperature control in cascade with the steam flowrate to the reboiler—shown for both the summer and winter cases—is detailed in TABLE 1.

Assumptions are based on the following:

- Feed/reflux ratio is maintained without change
- Condenser outlet temperature is reduced
- Reboiler and condenser duties are calculated
- Reflux minimum temperature is calculated at maximum duty of condenser or reboiler
- Product specifications shall be met for all cases.

where:

Condenser design duty (MMBtu/hr): 164.1

Reboiler design duty (MMBtu/hr): 186.9

TABLE 1. Bottom temperature control in cascade with the steam flowrate to the reboiler: summer case (top) and winter case (bottom)

Description	Case 1 at design flowrate	Case 2 at design flowrate	Case 3 at turndown
Summer Case			
Condenser duty, MMBtu/hr	142.1	161.6	107.3
Reboiler duty, MMBtu/hr	172.9	186.7	93.3
Reflux temperature, °F	140	120	60
Winter Case			
Condenser duty, MMBtu/hr	146	164.1	75.7
Reboiler duty, MMBtu/hr	172.9	178.9	77.7
Reflux temperature, °F	140	91	60

TABLE 2. Top temperature control in cascade with reflux: summer case (top) and winter case (bottom)

Description	Case 1 at design flowrate	Case 2 at design flowrate	Case 3 at turndown
Summer Case			
Condenser duty, MMBtu/hr	142.1	164.4	66.7
Reboiler duty, MMBtu/hr	172.9	172.9	69.2
Reflux temperature, °F	140	60	60
Winter Case			
Condenser duty, MMBtu/hr	145.6	164	66.7
Reboiler duty, MMBtu/hr	172.9	172.9	69.2
Reflux temperature, °F	140	66	60

TABLE 3. Minimum product temperatures at the air exchanger outlet based on analysis

	Bottom temperature control, °F		Top temperature control, °F	
	At design flowrate, summer/winter	At turndown, summer/winter	At design flowrate, summer/winter	At turndown, summer/winter
Depropanizer	120/91	60/60	60/66	60/60

TABLE 4. Day and night ambient temperature changes

Air temperature, °F	Winter	Summer
Day	64.4	120.2
Night	41	82.4

Product specifications: C_4 in the distillate ≤ 0.0150 vol% and C_3 in the bottom ≤ 0.0066 vol%
Feed/reflux ratio (wt): 1.088
Condenser meets the thermal and hydraulic requirements.

The top temperature control in cascade with the reflux—for both the summer and winter cases—is shown in **TABLE 2.**

The following assumptions and basis are used:

- Steam flowrate to the reboiler is maintained
- Reflux flowrate and temperature are changed until the design duty of the condenser or reboiler are achieved
- Reflux flowrate shall be higher than that managed at turndown condition
- Feed flowrate is maintained constant
- Product specifications shall be met for all cases.

where:

Condenser design duty (MMBtu/hr): 164.1
Reboiler design duty (MMBtu/hr): 186.9
Product specifications: C_4 in the distillate ≤ 0.0150 vol% and C_3 in the bottom ≤ 0.0066 vol%
Feed/reflux ratio (wt): 1.088
Condenser meets the thermal and hydraulic requirements.

TABLE 3 summarizes the minimum product temperatures at the air exchanger outlet from the above analysis.

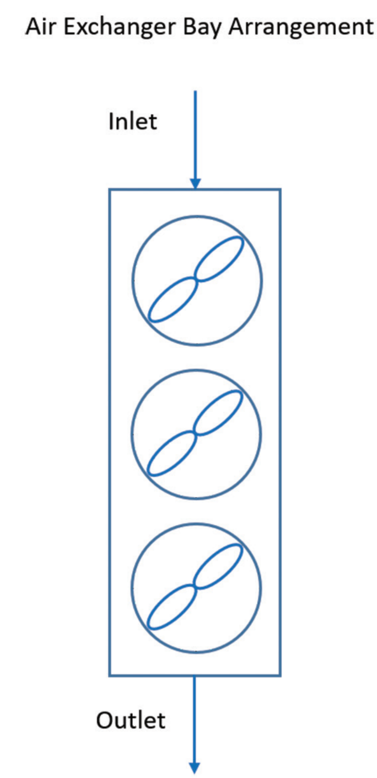


FIG. 2. Air exchanger bay arrangement.

Days/nights temperature variability vs. air exchanger operation/control.

Each air exchanger bay consists of three fans, as indicated in **FIG. 2.** The day and night ambient temperature changes naturally. The maximum product temperature at the condenser outlet (= 140°F) was analyzed with the given ambient conditions in **TABLE 4** to determine the required arrangement of variable-speed motors and step-control on-off.

The number of fans in operation [on-off fans and variable-frequency drive (VFD) fan)] have been validated at maximum flowrate (summer), minimum flowrate (winter) and turndown conditions (**FIG. 3**), using the heat transfer coefficients "Uclean" and "Udirty" for tem-

TABLE 5. Maintaining maximum product temperature at the air exchanger outlet (= 140°F)

Air temperature, °F	Service	Number of fans[1]	Turndown conditions
120	Using Uclean/product temperature, °F	2 + 40%/137	1 + 10%/128
110		1 + 60%/139	–
100		1 + 20%/139	–
90		1 + 10%/124	–
82.4		0 + 90/136	0 + 25%/135
120	Using Udirty/product temperature, °F	2 + 60%/139	1 + 10%/133
110		1 + 80%/139	–
100		1 + 30%/135	–
90		1 + 10%/136	–
82.4		1/140	0 + 25%/139
64.4	Using Uclean/product temperature, °F	0 + 44%/138	0 + 18%/136
54.4		0 + 36%/140	0 + 16%/131
41		0 + 30%/136	0 + 13%/136
64.4	Using Udirty/product temperature, °F	0 + 47%/139	0 + 18%/139
54.4		0 + 39%/137	0 + 16%/133
41		0 + 31%/138	0 + 13%/138

[1] The result representation = number of fans on full speed + % of duty required to be modulated/outlet air exchanger temperature

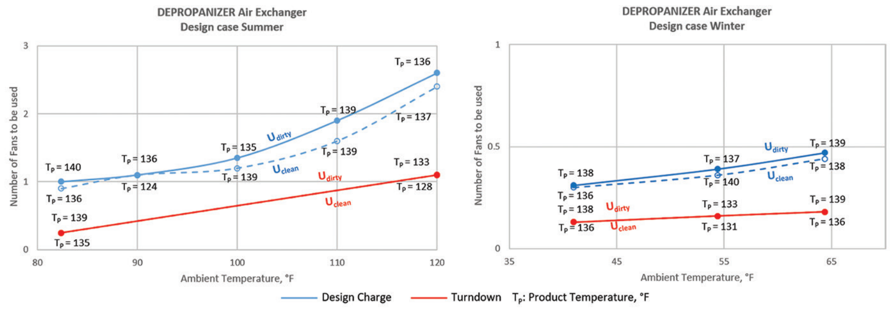
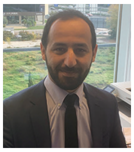


FIG. 3. Number of fans to be used during summer and winter cases.

perature intervals between day and night.

It has been proven that the maximum product temperature at the air exchanger outlet (= 140°F) can be maintained using a sequential control on-off in combination with one modulating fan, as summarized in **TABLE 5.**

Takeaway. Through the analysis described here, it was determined that each air exchanger bay (comprising three fans) can be configured with two fans, based on on-off control, while the third fan can be modulated through variable-speed motors. The analysis provides a means to establish the optimum selection of the control combination between variable-speed motors and on-off step controls. This minimizes capital costs in equipping all fans with variable-speed motors, and lowers operating costs by establishing the correct control method. **GP**



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LNG today: Meeting changing infrastructure requirements

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The growing global climate crisis is necessitating the search and need for cleaner energy choices. Changing market dynamics are driving evolution and growth within several sub segments such as LNG.

The need to meet sustainability goals in more mature global markets is accelerating infrastructure changes. As LNG applications evolve, facilities must adapt innovative infrastructure and vessels to ensure they continue to operate in a safe, efficient, cost-effective and timely manner. As the transition toward a net-zero economy gains momentum and LNG becomes increasingly cost competitive, the demand for LNG and its related infrastructure will only increase.

The evolution of LNG applications. Operators are improving infrastructure by upgrading jetties to support larger vessel types, using floating units as semi-permanent storage structures, with onboard liquid-to-gas conversion (e.g., FSRU) or using ship-to-ship transfers more regularly to meet demand. However, if paired with existing infrastructure that is no longer fit for purpose, all these potential solutions will impact efficiency and—more importantly—safety.

One-way foresighted ports and terminals are reducing disruption is by adding or modifying their facilities to offer LNG bunkering services gradually over time. Many of these shore-based facilities are in regions with tighter emissions control regulations and close to LNG import terminals for efficient distribution.

However, due to its low capital investment and the limited infrastructure required, truck-to-ship is one of the most widely used configuration at terminals and ports today for LNG. This method does have its drawbacks; among other factors, it requires multiple vehicles, ultimately limiting bunkering operations to smaller-sized LNG-fueled vessels.

Alternative options—such as ship-to-ship and terminal shore-to-ship transfers—support larger storage capacity and higher bunkering rates, but both methods require significantly higher capital investments for bunker vessels and fixed infrastructure, such as storage tanks, specialized loading arms and flexible hoses.

As LNG applications evolve, infrastructure and operations must evolve to be more flexible to avoid implications on efficiency and safety. For example, LNG bunker vessels must be able to both visit large terminals and provide transfers to LNG-fueled ships via ship-to-ship.

Infrastructure development and the LNG supply chain. LNG fueling is becoming an alternative for shipping lines wishing to reduce their carbon footprint with immediate impact. Using LNG helps lessen carbon dioxide (CO_2), nitric oxide (NO_x) and sulfur oxide (SO_x) emissions and has contributed to the development of new markets within the LNG industry, initiating unprecedented levels of ship and bunker vessel building and new gas train construction.

The new global limit of 0.5% of sulfur content in marine fuels—enforced by the International Maritime Organization (IMO) in January 2020—is poised to incentivize the investment in LNG. This stricter cap on marine bunker fuel is spurring the installation of new machinery (or conversion where possible) designed to operate on LNG, as well as the construction of related infrastructure to accommodate the switch to LNG-fueled vessels. This standard is creating a self-reinforcing feedback loop, where the development of an efficient, secure, and competitive LNG supply chain and related bunkering infrastructure drives further adoption of LNG-fueled vessels.

The LNG supply chain is a carbon-intensive process. Uptake in gas demand will be met by LNG in many countries without domestic gas production or pipeline gas from nearby countries. By its very nature, the LNG supply chain spans the globe and involves different industry processes. However, the emissions from LNG have been considered on a more

FIG. 1. View of a floating production storage and offloading (FPSO) vessel.

FIG. 2. LNG carrier discharging at a terminal. Photo courtesy of Trelleborg.

segmented basis. With the growth of the LNG fueling market, there is an increased focus on the lifecycle emissions of the whole LNG supply chain—from 'well-to-wake' emissions to final combustion.[1]

Projects attracting investment. LNG is a global commodity, with 21 countries exporting to 42 importers. The bunkering infrastructure to support LNG as a marine fuel continues to snowball.

Investment in LNG infrastructure has grown with 124 ports now providing LNG bunkering facilities.[2] In early 2019, there were just six LNG bunkering vessels in operation; five in Europe and one in North America. As of July 2020, the number has more than doubled, growing to 13 in service, with a further 28 on order and/or undergoing commissioning.[1]

With 120 LNG-powered ships in service around the world, and another 130 on order[3], an increasing number of bunkering facilities are gearing up to support the demand for LNG as a fuel into the future.

Flexible LNG solutions and improving interface management. Across international markets, LNG is traded as a commodity. In international shipping, it is used as a fuel. Each market requires flexible solutions to ensure safety, efficiency, cost effectiveness and, ultimately, the success of the business model—from ship-shore links for FSRUs to hybrid-GEN3 solutions for bunker vessels. New projects must find a fast return on investment (ROI), while established facilities must keep pace with changing demands.

Given the global scope and myriad applications of the LNG industry, diversity is the norm. From traditional terminals to bunker barges and everything in between, project requirements vary substantially, inviting varied solutions and complicating interface management at transfer touchpoints. Optimizing the interface at the various stages of the LNG supply chain is critical to supporting the business model of every transfer operation. Interface optimization means consistent communication and standardized processes at every transfer point.

Efficient equipment delivers flexibility. Adopting easily configurable and compatible equipment systems delivers several benefits, such as an enhanced overview of operations, improved productivity, reliability and safety, and a faster ROI for all stakeholders. Efficient systems that offer these benefits can provide support to LNG operators that require operational flexibility to adapt to spot contracts.

Conversely, fragmentation creates inefficiencies and safety issues and reduces the opportunity to implement flexible business models. A standardized approach across facilities opens opportunities for all stakeholders through common requirements and systems. Standardization of systems improves operational control. At the same time, data sharing between parties is enhanced, enabling effective communications, fast response to potential issues and empowered long-term decision-making.

To support an international shipping network fueled by LNG requires a robust system architecture design at the outset. In turn, this requires oversight between stakeholders and an understanding of cross-party requirements.

The role of specifications. Every port and terminal are unique. It is important to identify the correct specifications needed at the early stages of a project to ensure long-term performance and the safety of the project. The ability to understand materials and applications plays a large part in optimizing safety and performance and ensuring the right solution for the job.

At the same time, products must meet differing regulatory requirements globally, and suppliers must understand and integrate all necessary standards into their solution. They must be prepared to provide first-class 24/7 support when it is needed to ensure downtime is minimized.

Looking ahead. Demand for cleaner fuels is set to propel LNG fueling into a mature market phase, where spot contracts are utilized, rather than solely long-term contracts. In addition to developing economies driving new applications and markets, the global LNG infrastructure market is expected to witness significant growth in the near future.

However, LNG infrastructure must be able to keep up with demand. Accelerating LNG fueling to meet sustainability demands requires LNG infrastructure that can cope with demand by berthing more LNG-powered vessels safely and efficiently.

To respond to LNG's various challenges and opportunities, LNG leaders must adapt to the needs of different business models, changing environments and transfer scenarios. To do this and help ensure LNG operations take place safely and efficiently, operational flexibility is crucial. **GP**

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Case study: Seal-less caustic circulation pump failure

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A Saudi Aramco gas plant contains two (2 × 100%) lean caustic circulation seal-less pumps. The pumps are responsible for delivering the lean regenerated caustic with naphtha from the disulfides separator to the naphtha wash settler. The pumps are considered an essential component of the caustic regeneration loop that ensures removing the total sulfur from the natural gas liquids (NGL) products.

Pump design. The pumps are 178-hp seal-less magnetic drive pumps with a rated flow of 294 gal/min (gpm) and a total differential head of 638 ft. The process fluid is totally contained inside the pump containment shell and depended on to remove the generated heat through internal lubrication paths. Therefore, the slightest restriction to the lubrication flow paths will jeopardize the condition of the pump internal parts, resulting in a catastrophic failure. Additionally, the fluid runs between the inner magnet, the containment shell and through the shaft holes to the rear of the pump shaft, where it returns to the pump's suction through the thrust balance hole in the impeller (FIG. 1)

The pump shaft connected to the inner magnet is rotated by the action of the magnetic flux circuit between the inner and outer magnets. The outer magnet is coupled to the driver motor through another shaft to transmit the rotation to the inner magnet.

Background. Within normal operating conditions, the pump was tripped due to a motor thermal overload fault signal. All process conditions were checked and found to be within normal parameters. The pump was later started again and immediately tripped on a stalled motor fault signal.

It was decided to remove the pump for dismantling and internal inspection. Major findings included:

- Excessive rubbing occurred between the impeller front cover and the pump casing, as well as thinning on the impeller shrouds. This is an indication of high thrust movement.
- The rear thrust pad, thrust washer and sleeve bush were not found in place. The silicon carbide bearings were found to have completely failed and their traces were found inside the pump casing (FIG. 2).

Investigation. Based on the inspection findings listed above, operational trends, caustic lab sample analysis and metallurgical analysis, the following conclusions were reached:

- Material analysis showed that grains of the titanium sleeve bush were colored blue, consistent with the newly introduced process catalyst in 2019. This was considered a sign of a corrosion reaction. It is suspected that the catalyst filled the standby pump when it was not in operation—the pump would have been unable to disperse the liquid. The product would have then attacked the front sleeve bush, as this is mounted in the casing volute, whereas the rear sleeve bush is in a less accessible area of the pump.
- Occasionally, the process is subjected to high caustic concentration. Material analysis shows that there should be an accelerated rate of corrosion of the titanium sleeve bush carrier in instances where the pump is exposed to higher (40%–50%) caustic concentrations.

As shown in FIG. 3, it is suspected that corrosion and the loss of the titanium sleeve bush circumferential were enough to allow the bolts to loosen their grip between the bush carrier and holder. Consequently, all of the bush carriers became free to move radially and axially, leading to the thrust pad and thrust washer silicon carbide to shatter.

The failure of the rear thrust pad and thrust washer (active bearing) increased the negative thrusting and axial movement of the whole rotor towards the pump casing. This resulted in heavy wear on the casing, impeller and radial bush failure.

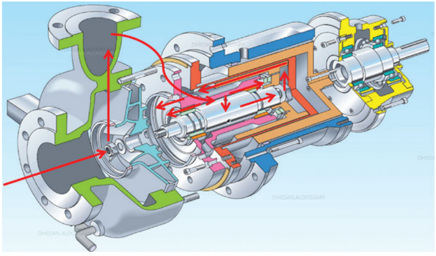

FIG. 1. Magnetic drive pump lubrication flow regime.

FIG. 2. Rear bush housing assembly (bush carrier) with damaged thrust pad corrosion/erosion impact. Additionally, signs of chemical attack (corrosion) were found in the front NDE titanium bush housing assembly.

Recommendation. It was concluded that the impact of the intermittent 50% caustic concentration and the new catalyst type corroded the titanium part, resulting in thrust bearing failure.

The vendor confirmed that the use of titanium is not a common selection, but the request had been made by the customer during the project design. Additionally, it was noticed that the other pump components that are made of duplex stainless steel did not experience any corrosion attack.

From these findings, vendor experience and the corrosion chart, it was recommended to upgrade the bush housing material upgrade to duplex stainless-steel material. **GP**

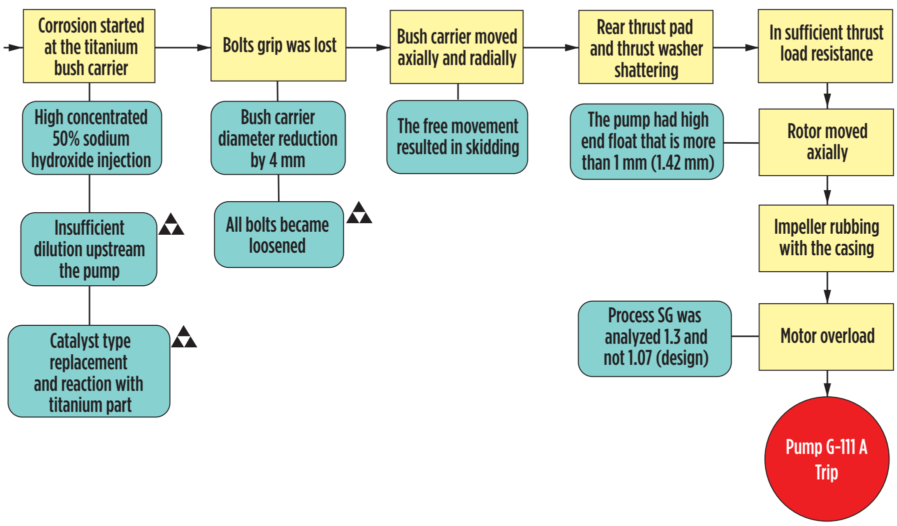

FIG. 3. Proprietary process[a] identifying the casual factors and sequence of a pump trip.

NOTES

[a] TapRoot

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Chesterton Connect equipment monitoring sensor for use in hazardous environments

Chesterton has released the equipment monitoring sensor, Chesterton Connect for use in hazardous conditions. This is part of Chesterton's IoT wireless monitoring product line, which is certified for use on equipment and structures that are in environments with high pressures, high temperatures and flammable liquids.

The Chesterton Connect is rated for Class 1/Division 1 (gas, vapor environments) and Class 2/ Division 1 (dust environments). The sensor contains electrical protection that prevents it from becoming a source of ignition.

The Connect sensor is capable of monitoring rotating equipment such as pumps and heat exchangers for equipment vibration, equipment surface temperature, process pressure and process temperature. The sensor communicates through Bluetooth and sends alerts to the user's app.

Drager released methane detection cameras

The MetCam optical gas detection camera, by Drager, automatically identifies methane leaks. This technology can prevent potentially hazardous gases from damaging the environment and disrupting the safety of the plant.

The camera displays a black and white video image—the gas gas is colored so that it can be easily detected—the camera quantifies the concentration of the escaping gas. In addition, this technology automatically recognizes when the optics are obscured or dirty to prevent false alarms. The MetCam self-calibrates on a set schedule and adapts to changing weather conditions.

Technip Energies, Svante to develop CO_2 capture projects in Europe and the Middle East

Technip Energies and Svante have entered an MOU to further develop Svante's solid sorbent carbon capture technology and provide integrated solutions from concept to project delivery.

The partnership will explore opportunities in Europe, the Middle East and Africa and Russian Federation markets where Svante's technology would be selected by end users for industrial carbon capture projects, including cement & limestone, blue hydrogen, refineries, petrochemicals, steel, ammonia and pulp & paper facilities. The cooperation will be worldwide for blue hydrogen plants using Technip Energies' steam methane reformer technology.

The carbon-capture facilities will use Svante's solid sorbent technology to capture carbon directly from industrial post-combustion flue gases as a non-intrusive "end-of-the-pipe" solution to produce pipeline-grade CO_2. Svante's net-zero technology captures CO_2, concentrates it, and releases it for safe storage or industrial use, all in less than 60 sec, by using proprietary active capture nanomaterials called solid sorbent filter.

The Lee Company has introduced new pumps for high-pressure, high-temperature applications

The Lee Company released a new positive displacement axial piston pump, which offers increased performance in high-pressure, high-temperature applications. The HPHT pumps family can generate up to 10,000 psid at temperatures of 400°F. This is ideal for critical hydraulic fluids transfer applications. The axial piston design does not contain any elastomers, which increases resistance and improves reliability.

HTHP pumps are designed to be driven by a Maxon Motor and are qualified for more than 1,000 hr of operation while withstanding high shock and vibration levels. The pumps are available in 22-mm and 30-mm diameter options. Cylindrical Hi-Bar inlet screens that slide onto the body of the pump are also available.

Sundyne's HMD Kontro launched complete fluid containment solution

HMD Kontro's latest magnetic drive seal-less pump solution offers safe transfer of hydrofluoric acid during the alkylation processes. The technology was developed in collaboration with hydrofluoric acid unit operators and Honeywell UOP. The HMD Kontro hydrofluoric acid pump is corrosion resistant, contains leak detection, protection against dry pump running and is safer for people and the environment.

Mechanical seals are challenged in alkylate applications and can leak harmful process fluids. Designed for resistance against highly corrosive fluids, the HMD Kontro application-specific, seal-less pump ensures conformity with all environmental health and safety standards, protecting people, plant and the environment through complete fluid containment.

MOL, Vopak to build Hong Kong's first LNG import terminal

Mitsui O.S.K. Lines Ltd. (MOL) and Vopak announced an agreement has been reached, whereby Vopak will acquire 49.99% of the shares in the vessel owning company of MOL FSRU *Challenger,* whose name will be changed to *Bauhinia Spirit.* This new JV company between MOL and Vopak in Hong Kong will own the world's largest floating storage and regasification unit (FSRU) and have a long-term contract with Hong Kong LNG Terminal Ltd. The FSRU has a storage and regasification capacity of 263,000 m^3 and 800 $MMsft^3d$, respectively. Under the contract, the JV will provide the FSRU as well as jetty operations and maintenance and port services.

The offshore jetty platform for the mooring of the FSRU and LNG carriers are owned by Hong Kong LNG Terminal Ltd.

The terminal is under construction and is expected to be operational around mid-2022. The terminal will be located offshore about 25 km southwest of Hong Kong Island and will provide natural gas feedstock to the customer's dedicated power plants. It is being developed to support the Hong Kong Special Administrative Region government's target to improve air quality and environmental conditions by increasing the percentage of power generation by natural gas.

TotalEnergies and Clean Energy launched the construction of their first biogas unit

TotalEnergies and its U.S. partner Clean Energy launched the construction of their first biomethane production unit, in Friona, Texas. Biomethane will be used as an alternative fuel for mobility, thus contributing to decarbonize road transportation.

Located on the Del Rio dairy farm, the facility will be fueled by the onsite supply of livestock manure to produce more than 40 GWh of biomethane per year. The biomethane will be distributed in the U.S. by Clean Energy through its network of fueling stations, enabling the supply of renewable gas to between 200 and 300 trucks per year.

By processing cow manure, a significant source of methane emissions, and substituting fossil fuels with renewable energies, the project will avoid some 45,000 tpy of CO_2e emissions.

FCI launched technology to measure methane

Fluid Components International (FCI) debuted the ST80 Series Thermal Mass Flow Meter to measure methane and provide accurate emissions data for companies to meet environmental regulations and reporting requirements.

The ST80 Series uses hybrid sensor drive adaptive sensor technology (AST). This measuring technique combines constant power and constant temperature thermal dispersion sensing technologies. The meters have also been approved internationally for use in Div.1/Zone 1 environments.

The ST80 offers a choice of four precision flow sensor element designs, one of which is the Wet Gas MASSter sensor developed for the ST80 Series. The Wet Gas MASSter optimizes the sensor head design and installation to prevent condensation droplets, entrained moisture or rain from contacting the thermowells.

The meters are suitable for pipe diameters from 1 in.–99 in. and air/gas temperatures up to 850°F. They feature an accuracy of ±1% of reading, ±0.5% of full scale and repeatability of ±0.5% of reading with flowrates ranging from 0.2 ft/sec–1,000 ft/sec (0.07 m/sec–305 m/sec) and 100:1 turndown.

ExxonMobil, Scepter Inc. to deploy satellite technology for real-time methane emissions detection

ExxonMobil and Scepter, Inc. have agreed to work together to deploy advanced satellite technology and proprietary data processing platforms to detect methane emissions at a global scale. The agreement has the potential to redefine methane detection and mitigation efforts and could contribute to broader satellite-based emissions reduction efforts across a dozen industries, including energy, agriculture, manufacturing and transportation.

In the first phase of the project, the companies will design and optimize the plan for satellite placement and coverage, initially focusing on capturing methane emissions data from ExxonMobil operations in the Permian Basin. Scepter will deploy satellites in 2023 and increase coverage to more than 24 satellites over 3 yr, forming a large constellation network capable of monitoring operations around the world.

Scepter's satellite detection technology has shown the ability to accurately collect data on methane, while also identifying sources of CO_2, nitrogen oxides, sulfur oxides and other greenhouse gases. When combined with ExxonMobil's data from ground-based sensors and aerial surveys using advanced analytics, Scepter's data platform allows the company to further establish information regarding its methane emissions performance.